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EVALUATION OF POLLUTION CONTROL IN FOSSIL FUEL CONVERSION PROCESSES. LIQUEFACTION: SECTION I. COED PROCESS

C. D. Kalfadelis, et al

Exxon Research and Engineering Company

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The report gives results of a review of the FMC Corporation's COED coal conversion process, from the standpoint of its potential for affecting the environment. It includes estimates of the quantities of solid, liquid, and gaseous effluents, where possible, as well as the thermal efficiency of the process. It proposes a number of possible process modifications or alternatives, and points out new technology needs, aimed at lessening adverse environmental impact.

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LIQUEFACTION: SECTION I. COED PROCESS

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SUMMARY

The FMC Corporation's COED coal conversion process has been reviewed from the standpoint of its potential for affecting the environment. The quantities of solid, liquid and gaseous effluents have been estimated, where possible, as well as the thermal efficiency of the process. A number of possible process modifications or alternatives have been proposed and new technology needs have been cited, with the main objective the lessening of adverse environmental impact.

TABLE OF CONVERSION UNITS

To Convert From	To	Multiply By
Btu	Calories, kg	0.25198
Btu/pound	Calories, kg/kilogram	0.55552
Cubic feet/day	Cubic meters/day	0.028317
Feet	Meters	0.30480
Gallons/minute	Cubic meters/minute	0.0037854
Inches	Centimeters	2.5400
Pounds	Kilograms	0.45359
Pounds/Btu	Kilograms/calorie,kg	1.8001
Pounds/hour	Kilograms/hour	0.45359
Pounds/square inch	Kilograms/square centimeter	0.070307
Tons	Metric tons	0.90719
Tons/day	Metric tons/day	0.90719

INTRODUCTION

Along with improved control of air and water pollution, the country is faced with urgent needs for energy sources. To improve the energy situation, intensive efforts are under way to upgrade coal, the most plentiful domestic fuel, to liquid and gaseous fuels which give less pollution. Other processes are intended to convert liquid fuels to gas. A few of the coal gasification processes are already commerically proven, and several others are being developed in large pilot plants. These programs are extensive and will cost millions of dollars, but this is warranted by the projected high cost for commercial gasification plants and the wide application expected in order to meet national needs. Coal conversion is faced with potential pollution problems that are common to coal-burning electric utility power plants in addition to pollution problems peculiar to the conversion process. It is thus important to examine the alternate conversion processes from the standpoint of pollution and thermal efficiencies and these should be compared with direct coal utilization when applicable. This type of examination is needed well before plans are initiated for commercial applications. Therefore, the Environmental Protection Agency arranged for such a study to be made by Exxon (formerly Esso) Research & Engineering Company under contract EPA-68-02-0629, using all available non-proprietary information.

The present study, under the contract, involves preliminary design work to assure the processes are free from pollution where pollution abatement techniques are available, to determine the overall efficiency of the processes and to point out areas where present technology and information are not available to assure that the processes are non-polluting.

All significant input streams to the processes must be defined, as well as all effluents and their compositions. This requires complete mass and energy balances to define all gas, liquid, and solid streams. With this information, facilities for control of pollution can be examined and modified as required to meet Environmental Protection Agency objectives. Thermal efficiency is also calculated, since it indicates the amount of waste heat that must be rejected to ambient air and water and is related to the total pollution necessary to produce a given quantity of clean fuel. Alternatively, it is a way of estimating the amount of raw fuel resources that is consumed in making the relatively pollution-free fuel. At this time of energy shortage this is an important consideration. Suggestions are included concerning technology gaps that exist for techniques to control pollution or conserve energy. Maximum use was made of the literature and information available from developers. Visits with some of the developers were made, when it appeared warranted, to develop and update published information. Not included in this study are such areas as cost, economics, operability, etc. Coal mining and general offsite facilities are not within the scope of this study.

Considerable assistance was received in making this study, and we wish to acknowledge the help and information furnished by EPA and FMC Corporation.

1. PROCESS - GENERAL

1.1 Process History

The COED process has been under development by FMC Corporation as Project COED (Char-Oil-Energy Development) since 1962 under the sponsorship of the Office of Coal Research of the U.S. Department of the Interior (1-12). Bench-scale experiments led the way to design and construction in 1965 of a process development unit (PDU) employing multi-stage fluidized-bed pyrolysis to process 50-100 pounds of coal per hour (1,13). Work with the PDU was extended to other coals in 1966, and hydrotreating of COED oil from the PDU was studied by Atlantic Richfield Company (2). Correlated studies included an investigation of char-oil and char-water slurry pipelining economics, high-temperature hydrogenation for char desulfurization, and an economic appraisal of the value of synthetic crude oil produced from COED oil.

In a second contract phase, additional coals were processed and COED economics were updated to 1970 (3). The COED char desulfurization effort was concluded with a recommendation to explore char gasification alternatively (4). And a COED pilot-plant processing 36 TPD of coal and able to hydrotreat 30 BPD of oil was designed and constructed in 1970 (5).

The pilot plant was operated successfully on a number of coals under terms of a third contract (6) in 1971-72. This contract phase also extended oil filtration studies on a rotary pressure filter; and an oil absorber tower was designed to replace the aqueous condensation system in the product recovery train. The oil absorption system was intended to reduce or eliminate the filtration of pyrolysis oil. The system was being installed in June, 1972.

The American Oil Company prepared an independent economic evaluation of the COED process in 1972 (7). In this case, char was to be gasified using the Kellogg molten salt process at low pressure in order to conserve the sensible heat of hot char.

Development of the COED process is continuing, with major funding provided by OCR (14). The character of the process has changed in the course of development, and, even now, it is difficult to characterize it completely. This is due largely to the process variants which may be applied to treat product char, which may represent 50-60 weight percent of the coal fed to the process, and to the possible variants relating to gas treatment and to the supply of fuel to the process. What has remained constant in the development is the use of multi-staged fluidized beds operated at low pressure and at successively higher temperatures to pyrolyze a variety of high-volatile bituminous and semi-bituminous coals continuously. The achievements of the development program in this area of fluidized bed technology, albeit on a relatively small physical scale, are significant.

The process basis for our evaluation is the design study developed by FMC Corporation in 1973 for a "25,000 tpd COED plant" (15). Process flowsheets were developed for the pyrolysis plant, raw oil filtration section, and for the hydrotreating facility (Appendix Figures 1-5). A block flow diagram of these processing sections is shown in Figure 1. This design feeds 25,512 tpd of an Illinois No. 6-seam coal containing 5.9% moisture, 10.6% ash, and 3.8% sulfur. 12,512 tpd of product char is recovered, along with 3945 tpd of hydrotreated oil (24,925 bpd of indicated 25° API gravity). Flowsheets were not developed for coal preparation, gas treatment, hydrogen manufacture, oxygen manufacture, sulfur production, water and waste treatment, or utilities generation.

In the vast body of information which has been published relative to the pilot plant work (1-19), there is very little so far which relates to the pollution potential of the COED process. The thrust of the work has been directed to process development, including hardware development and yield improvements. A recent paper (30) does summarize an integrated scheme whereby pollution may be held to low levels. Flue-gas treatment is avoided in the main processing sections by firing clean product gas in dryers and heaters. Aqueous contaminated process condensates are stripped of H_2S and NH_3 with product gas and recycled to the last pyrolyzer (or are directed to the char gasifier, if the plant includes char gasification), where organic contaminants may be consumed. H_2S and NH_3 are removed from product gas by commercial processes, and sulfur and ammonia are sold.

The pollution potential has not been completely defined in the context of U.S. standards even for those coal gasification processes which have already been commercialized elsewhere. Standards have changed radically in recent years, and new standards continue to be promulgated by governing agencies. Coal compositions, including sulfur and trace element contents, vary widely (52), and stream compositions from a particular process are generally sensitive to coal composition. Hence, although COED does produce a low-sulfur char from Utah A-seam coal, the coal itself is sufficiently low in sulfur to permit its use directly. Neither is the case with an Illinois No. 6-seam coal.

Future FMC research programs may be directed to a more precise determination of stream compositions relative to contaminant levels and to the effects of extended recycle of recovered contaminated liquors to pyrolysis (14).

1.2 Process Description

The COED process being developed by the FMC Corporation is a continuous, staged fluidized-bed coal pyrolysis operating at low pressure, and is designed to recover liquid, gaseous, and solid fuel components from the pyrolysis train. Heat for the pyrolysis is generated

by the reaction of oxygen with a portion of the char in the last pyrolysis stage, and is carried counter-currently through the train by the circulation of hot gases and char. Heat is also introduced by the air combustion of the gas used to dry feed coal and to heat fluidizing gas for the first stage. The number of stages in the pyrolysis and the operating temperatures in each may be varied to accomodate feed coals with widely ranging caking or agglomerating tendencies.

Oil that is condensed from the released volatiles is filtered on a rotary precoat pressure filter and catalytically hydrotreated at high pressure to produce a synthetic crude oil. Medium-Btu gas produced after the removal of acid gases is suitable as clean fuel, or may be converted to hydrogen or to high-Btu gas in auxiliary facilities. Residual char (50-60% of feed coal) that is produced has heating value and sulfur content about the same as feed coal, so that its ultimate utilization may largely determine process viability.

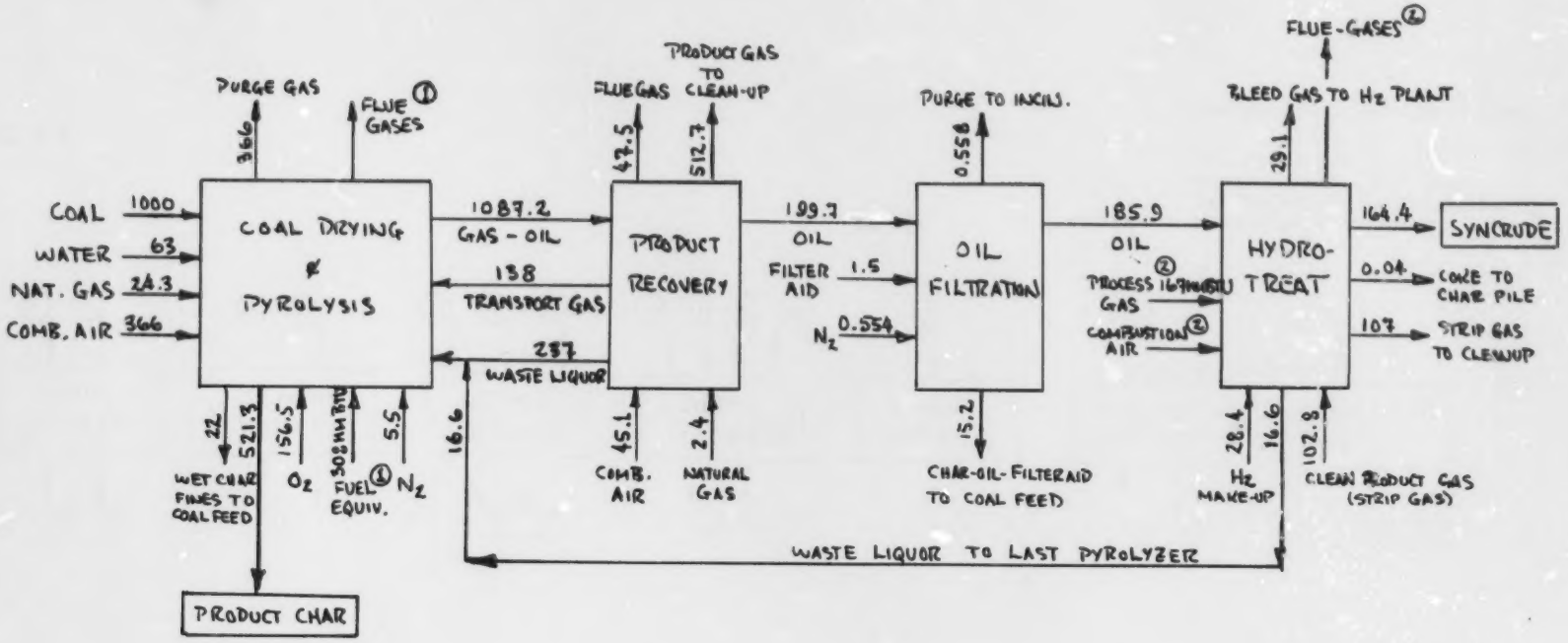

FIGURE 1 - COED COAL CONVERSION (15)

(All rates in tph)

1. Heating media not specified for oxygen heaters and steam superheaters; flue gas rates depend on fuels consumed.

2. Analysis of gas fired in hydrotreater preheater not specified; combustion air requirement and flue-gas rate depend on fuel consumed.

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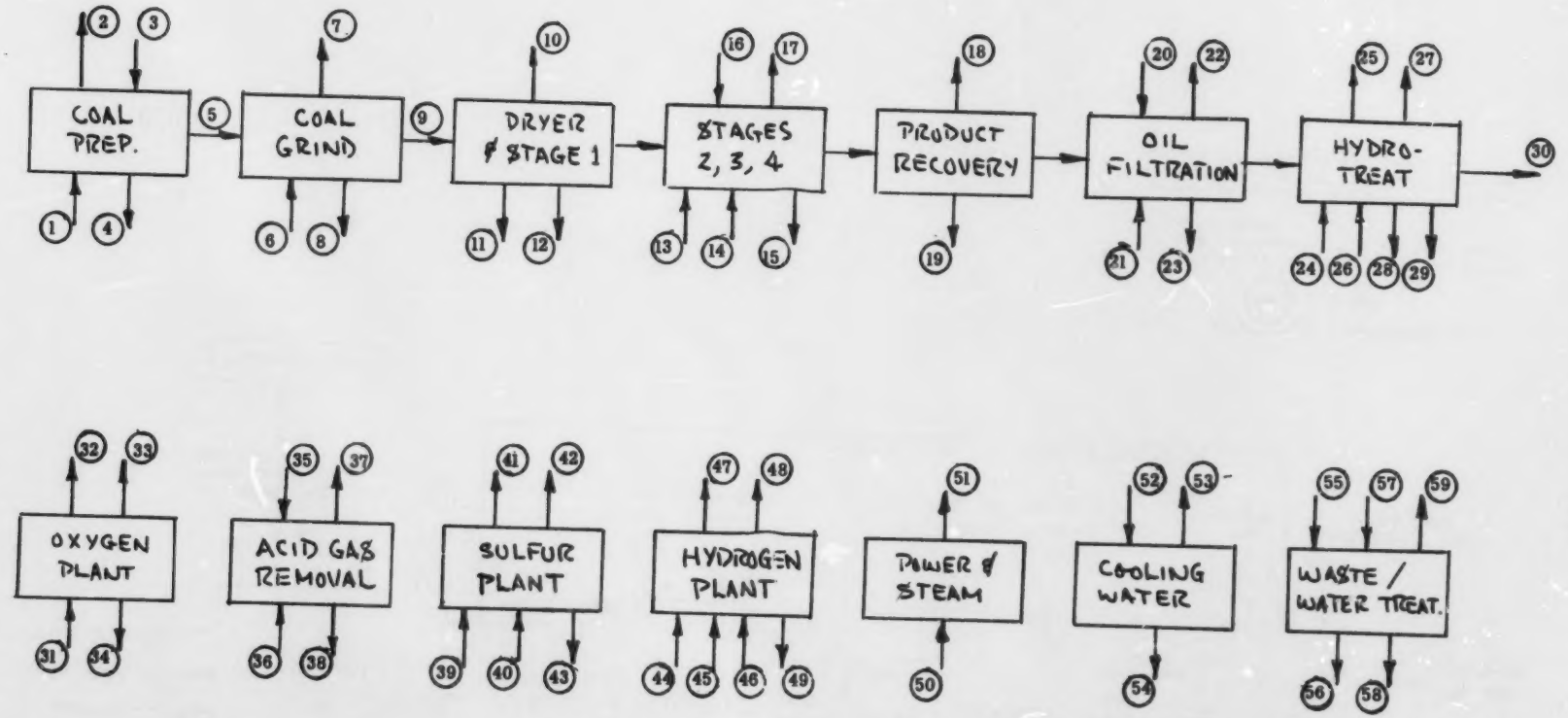

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FIGURE 2 - COED DESIGN REVISED TO INCORPORATE ENVIRONMENTAL CONTROLS AND TO INCLUDE AUXILIARY FACILITIES.

(See Table 1 for Stream Identification)

Table 1

Stream Identifications for Revised COED Process

(Stream Numbers Refer to Figure 2. See text for details)

Coal Preparation

1. Influence of weather (wind, temperature, humidity) on 40-50 acre on-site coal storage piles.

2. Dusting and wind losses; possible odor.

3. Precipitation on 40-50 acre storage area.

4. Storm run-off estimated at 10,000 gpm contains particulates and may be sulfidic. Directed to oily-water retention ponds along with run-off from processing areas for subsequent addition to waste water treatment system.

5. 1237 tph Illinois No. 6 seam coal, 14 percent moisture.

Coal Grinding

6. Approximately 455 MM Btu per hour input to dry coal (from 14 to 5.9 percent moisture). Boiler flue-gas stream may supply part of requirement.

7. 108 tph water removed from coal. Vent gas stream issuing through bag filters may require treatment to limit CO content.

8. 66 tph coal fines, 4 percent moisture, issues as fuel product.

9. 1063 tph sized coal, 5.9 percent moisture.

Coal Dryer and Stage 1 Pyrolysis

10. 366 tph purge gas requires treatment to limit CO content, directed to boiler stack.

11. 22 tph oily wet char fines separated at fines filter directed to coal feed.

12. 93.5 tph aqueous condensate. 83.3 tph directed to last pyrolyzer, 10.2 tph directed to water treatment.

Stage 2,3,4 Pyrolysis

13. 156.5 tph oxygen from oxygen plant.

14. 337 tph recycled process liquors as steam to last pyrolyzer.

15. 521 tph product char stream.

Table 1 (Cont'd)

Stream Identifications for Revised COED Process

(Stream Numbers Refer to Figure 2. See text for details)

16. 530 gpm BFW to fluidized bed char cooler and 360 gpm BFW to char after-cooler.

17. 265,000 lb/hr 600 psia steam and 180,000 lb/hr 150 psia steam from char cooling.

Product Recovery

18. 512.7 tph product gas to acid-gas removal system.

19. 236.7 tph aqueous condensate recycled to last pyrolyzer.

COED Oil Filtration

20. 1.5 tph (equivalent) filter aid supplied during filter precoat cycle; basecoat may also be used.

21. 0.5 tph nitrogen from oxygen plant filter pressurizing medium.

22. 0.5 tph purge gas directed to incinerator or boiler stack.

23. 15.2 tph oily char fines removed at filters containing 1.5 tph filter aid, recycled to coal feed.

Hydrotreating

24. 28.4 tph hydrogen make-up stream from hydrogen plant.

25. 29 tph bleed gas stream directed to clean-up and hydrogen plant for reprocessing.

26. 103 tph clean product gas used as stripping medium.

27. 107 tph contaminated gas directed to acid-gas removal system.

28. 0.04 tph reactor coke directed to product char pile. Spent catalysts require special treatment.

29. 16.6 tph contaminated aqueous condensate directed to last pyrolyzer.

30. 164.4 tph syncrude product.

Oxygen Plant

31. 440 MM scfd air intake.

32. 340 MM scfd nitrogen and other air constituents.

Table 1 (Cont'd)

Stream Identifications for Revised COED Process

(Stream Numbers Refer to Figure 2. See text for details)

33. 156.5 tph oxygen to pyrolysis.

34. 17 gpm water condensate from inter-coolers directed to boiler feedwater treatment.

Acid-Gas Removal

35. 512.7 tph product gas from pyrolysis.

36. 381,000 lb/hr 150 psig steam to regenerators.

37. 300 tph CO_2 and 14.4 tph H_2S directed to sulfur plant.

38. Spent Benfield solution and/or blowdown requires special treatment.

Sulfur Plant

39. 23 tph H_2S in incoming acid-gas streams.

40. 0.7 MM scfd regeneration air stream to Stretford solution in Beavon tail gas treatment.

41. Regeneration air stream directed to incinerator or to boiler stack. Stretford solution blowdown requires special treatment.

42. 150 MM scfd CO_2 containing less than 200 ppm sulfur.

43. 510 tpd elemental sulfur product.

Hydrogen Plant

44. 25 tph clean product gas feed to reformers.

45. 29 tph bleed stream from hydrotreating fed to reformers after clean-up.

46. 43 tph net water consumption in reformers (230 gpm BFW to reformers).

47. 60 tph CO_2 removed from reformer effluent.

Table 1 (Cont'd)

Stream Identifications for Revised COED Process

(Stream Numbers Refer to Figure 2. See text for details)

48. 28.4 tph product hydrogen stream to hydrotreating.

49. Spent catalysts and blowdown from acid-gas removal steps require special treatment.

Power-Steam Generation

50. 2032 MM Btu/hr fuel equivalent (see Section 3.5.1).

51. Flue-gases require desulfurization if char is fired to supply fuel shortfall. 6.4 tph ash generated if char is fired; returned to mine for burial.

Cooling Water

52. Chemical additives may include chromium or zinc compounds, acids, chlorine, phosphates, phenols, copper complexes; 9000 GPM make-up.

53. 6000 gpm water evaporated and 600 gpm drift loss.

54. 2400 gpm draw-off from cooling towers. May require special treatment before injection into waste-water treatment system.

Water Treatment

55. 2990 gpm total process water and 7590 gpm raw water make-up for treatment.

56. 10,580 gpm to users.

57. Additives to system may include lime, anti-foam, acids, char or activated carbon, oxygen or ozone and other agents.

58. Miscellaneous sludges from aeration, biox, and separation facilities may require special treatment.

59. Control of noxious evaporative losses may require special engineering, including floating covers on retention ponds or tanks and/or forced draft systems.

2. EFFLUENTS TO AIR - MAIN PROCESSING SECTIONS

All effluents to the air are shown in Figure 2 and Table 1. These effluents are based on the FMC Corporation design (Ref. 15 and Appendix Figures 1-5) and are to some degree inferred by analogy with prior art.

2.1 Coal Preparation and Storage

Common to all fuel coal usage, and particularly to coal conversion processes, are the operations of coal mining, which may include coal laundering, drying, and screening, coal transport, and storage. This study does not include energy and/or pollution considerations relative to these operations.

On-site coal storage will be required for all conversion plants to provide back-up for continuous conversion operations. For thirty days storage, there might be eight piles, each about 200 feet wide, 20 feet high, and 1000 feet long. Containment of air-borne dusts is generally the only air pollution control required for transport and storage operations, although odor may be a problem in some instances. Covered or enclosed conveyances with dust removal equipment may be necessary, but precautions must be taken against fire or explosion. Circulating gas streams which may be used to inert or blanket a particular operation or which may issue from drying operations will generally require treatment to limit particulate content before discharge to the atmosphere. Careful management and planning will minimize dusting and wind loss and the hazard of combustion in storage facilities.

The as-received feed coal employed in this design is indicated to have 10-14 weight percent moisture content. The FMC process basis feeds coal of about 5.9 weight percent moisture to the coal dryer ahead of the first pyrolyzer. Hence the free or surface moisture is assumed to be removed in the upstream coal preparation plant, although, obviously, the coal dryer proper may be arranged to remove a larger fraction of the original moisture.

We note that Illinois No. 6 coal is currently being supplied with about 17 percent moisture, but that this moisture content is a function of the operation of laundering equipment. In a commercial conversion plant situated at the mine, closer control of the delivered moisture would be possible, but with corresponding increase in energy consumption.

We note also that the reactivity of coals may be markedly affected by exposure to air, and that water serves to seal available pore volume, retarding oxidation. Hence the desired moisture content may be related to the average time-in-storage in a particular facility.

2.2 Coal Grinding

We have assumed that free moisture would be removed from feed coal by milling in a stream of hot combustion gases, as is practiced in the FMC pilot plant (5). Coal sized 16 Tyler mesh or smaller, but with minimum fines, is required for the pilot plant, although other studies (7) have indicated that particles up to 1/8 inch or 6-mesh may be suitable. In either case, the mechanical size reduction of an Illinois coal is expected to generate a considerable quantity of -200 mesh fines, especially if appreciable drying accompanies the milling operation. The quantity of such fines has been estimated to be 5 to 8 percent of the feed, depending on the type of equipment that may be used and on the acceptable size range, screening or separation efficiencies, and the recycle rates employed around the mill. Some small fraction of these fines will pass through the system with the sized coal. Additional fines will be produced in the coal dryer proper, and the ultimate consideration is that the total fines fed to the dryer or to the first pyrolyzer shall not overload the cyclone systems provided to effect their separation from the respective effluent streams. There may also be a relationship between the coal size fed to the system and the observable filter rates on raw pyrolysis oil. The fineness of char particles in Illinois No. 6-seam oils apparently contributed to blinding of the filter precoat in the pilot-plant filter (6). We have, therefore, assumed that fines generated in coal preparation, amounting to 5 percent of feed coal, will not be charged to pyrolysis, but will issue as a fuel product. Coal fines would probably be charged to the char gasification system, if this facility is included.

We have assumed that clean product gas is fired in the mill heater (the basis indicates that natural gas is used; see Section 5). About 110 tph of water must be removed if coal is received with 14 percent moisture. This may require the firing of 15-20 tph of product gas with 180-200 tph of combustion air in the milling circuit. Assuming a dry particulate separation system is adequate, bag filters might be used to recover fines from the vented gas following primary classification in cyclones.

Depending on water-use constraints, it may be desirable to condense water from the vent gas for reuse. This stream could be combined with, or treated similarly to, gas issuing from the coal drying and first-stage pyrolysis section, wherein the gas is scrubbed in venturi scrubber-coolers. The additional cooling requirement would be about equal to that provided in the design basis for treating vent gas from that section. It is presumed, however, that the additional coal fines separated from scrubber effluent by filtration in this way could not be recycled to the pyrolyzer, and would issue from the system as sludge. This sludge, containing 50 percent water, would preferentially be charged along with char to gasification, if char gasification is included, or might be combusted with char in a char boiler. However, the dry separation system employing bag filters would be preferred in the latter case.

Vent gas which issues from the bag filters from the milling circuit may contain a significant carbon monoxide concentration, depending on the combustion parameters employed in the mill. It may be necessary to direct the stream to a boiler stack or incinerator to complete the combustion. Another possibility is to employ a noble-metal catalytic afterburner (29), which would minimize the additional fuel requirement, to neutralize the stream.

2.3 Coal Drying and First Stage Pyrolysis

In FMC's design, clean natural gas is burned sub-stoichiometrically both to dry feed coal and to heat fluidizing gas for the first stage of pyrolysis. Both gas and air feeds to the heaters must be raised in pressure to match the operating pressures of the coal dryer and first stage, nominally 7-8 psig.

Coal is fed from storage hoppers by mechanical feeders into a mixing tee from which it is blown into the dryer with heated transport (recirculated) gas.

A cascade of two internal gas cyclones is provided both the coal dryer and the first pyrolysis reactor. Gas which issues from the first pyrolyzer is circulated through the fluidizing-gas heater for the coal dryer. Gas which issues from the coal dryer passes through an external cyclone and is then scrubbed in venturi scrubber-coolers, which serve to complete the removal of coal and char fines, as well as traces of coal liquids from the gas stream. Fines which are recovered in the external cyclone are passed through a mechanical feeder to a mixing tee where they are injected into the first-stage pyrolyzer by recirculated gas. Water equivalent to that introduced with coal and formed in the combustion processes is condensed from the gas in the scrubbing process.

Scrubber effluent passes into a gas-liquid separator, and the liquor stream is decanted and filtered to remove solids. The solids removed by filtration are indicated to amount to about one percent of the coal feed, and the wet filter cake is indicated to be recycled back to coal feed. The decanted liquor, except for a purge stream which, along with the filtrate from the fines filter, balances the removal of water from the section, is pumped back to the venturi scrubbers through water-cooled heat exchangers.

The gas stream which issues from the separator, except for a purge stream which removes the nitrogen introduced in the combustion processes, is compressed and recirculated to the gas heaters. This purge gas stream is essentially the only gaseous release from this section. Like the gas stream envisioned for the coal preparation section (see above), it is indicated to contain about 3.7 percent carbon monoxide, and will probably require further treatment before it may be released to the atmosphere. It may be possible to inject it into a boiler stack(s) along with air or oxygen to reduce CO emission. Alternatively the stream(s) may have to be incinerated in specific equipment for this purpose with additional fuel. The gas stream in this case represents a loss of combustible equivalent to about 230 MM Btu/hr. It is indicated to be sulfur-free (6, 14).

2.4 Stages 2,3,4 Pyrolysis

Coal which has undergone first-stage pyrolysis (at temperatures of about 550-600°F) is passed out of the stage into a mixing tee, from which it is transported into the second stage by heated recycle gas. Pyrolysis stages 2,3, and 4 are cascaded such that pyrolyzed solids pass through the stages in sequence in transport gas streams. Superheated steam and oxygen are injected into the last stage, where heat is released by partial combustion. Substantial recycle of hot (~1550°F) char from this last stage is used to supply heat to stages 2 and 3, in which it otherwise serves as an inert diluent. Similarly, hot gas which issues from the last stage is passed counter-currently through the cascade, serving also as the primary fluidizing medium in these reactors. Stages 2 and 3 operate at about 850° and 1050°F respectively.

The pyrolyzer vessels are each about 60-70 feet in diameter. A total of eight pyrolyzers in two trains is required to process the indicated feed coal. All fluidized vessels are equipped with internal dual-cascade cyclone systems.

Gas which issues from the second pyrolyzer passes through an external cyclone before being directed to the product recovery system. Fines which are separated are directed, along with product char from the last stage, to a fluidized bed cooler, which is used to generate 265,000 lb/hr. of 600 psia steam. First-stage recycle gas is used to fluidize the char cooler, and the gas which issues from the cooler is directed back to the venturi scrubbers in the first section after it has passed through an external cyclone. Fines from this cyclone are added to the char make from the last stage. Product char is available at this point at 800°F.

The FMC design indicates that char will be further cooled by cold-water exchange in unspecified equipment. In the pilot plant, a two-pass screw conveyor, in which cooling water is supplied to a hollow screw, as well as to the jackets of both flights, is used to cool char to about 100°F. About 180,000 lb/hr of 150 psia steam may be generated in the commercial operation if suitable equipment can be designed.

Because the system is otherwise closed, the only possible major atmospheric effluents from this section are the products of combustion from the heaters used to superheat the steam and oxygen feeds to the last pyrolysis stage. We have assumed clean product gas for this service also. About 10.5 tons of gas is required, along with about 105 tons of air per hour. The combustion products should be dischargeable directly in this case without further treatment.

2.5 Product Recovery System

Gas from the pyrolysis section is cooled and washed in two cascaded venturi scrubber stages to condense oil and solid components from the gas stream. The gas which issues from the second scrubber gas-liquid separator is passed through an electrostatic precipitator to remove microscopic

droplets, and is then cooled to 110°F by cold-water exchange to condense water. About a quarter of the gas stream is compressed and reheated for use as transport gas in the pyrolysis train. The remainder issues from the system as raw product gas, which is to be directed to an acid-gas removal system.

The oil and water condensed from the gas stream in the scrubber-coolers is decanted and separates into three phases: a light oil phase, a middle (aqueous phase), and a heavy oil phase. The oil phases are collected separately for dehydration in steam-jacketed vessels. The combined dehydrated oil is pumped to the COED oil filtration system.

A recycle liquor pump takes suction from the middle phase in the decanter. Recycle liquor is cooled in cold-water exchangers before being injected into the venturi scrubbers. Water condensed from the incoming gas leaves the section as a purge ahead of the recycle liquor coolers, and is indicated to be recirculated to the last pyrolysis stage.

The only major effluents to the atmosphere from this section are the combustion gases from the recycle transport-gas heater. Since clean product gas is fired in this heater, the combustion gases are dischargeable directly.

Vents from the oil decanters and dehydrators are indicated to be directed to an incinerator. Under normal operation, and with adequate condensing capacity in the vapor take-offs from the dehydrators, vent flow should be minimal.

2.6 COED Oil Filtration

FMC has designed a filtration plant to handle the COED raw oil output based on filtration rates demonstrated in its pilot plant (5,6). The system employs ten 700 $ft.^2$-rotary pressure precoat filters to remove char fines from the raw oil ahead of hydrotreating. Each filter is operated on a 7-hour precoat cycle, followed by a 41-hour filtration cycle.

Both the precoat and the raw oil to filtration are heated, using steam, to about 340°F. Inert gas (nitrogen) is compressed, heated, and recirculated for pressurizing the filters. The gas purge from the system, equivalent to the nitrogen make-up, is directed to an incinerator. It is indicated to contain only trace quantities of combustibles and sulfur.

Hot filter cake (38% oil, 52% char, 10% filter aid at 350°F) is discharged at the rate of about 15 tph, and is indicated to be added to the plant's char output in the process basis. FMC has recently indicated that filter cake will instead be recycled to coal feed (14). Filtered oil is directed to the hydrotreating facility.

2.7 Hydrotreating

Hydrotreating is employed to upgrade the heavy pyrolysis oil through the addition of hydrogen, which serves to convert sulfur to hydrogen sulfide, nitrogen to ammonia, and oxygen to water, as well as to increase the oil's hydrogen content through saturation reactions. Hydrotreating is performed catalytically in the pilot plant at 750 to 800°F and at total pressures of 2000-3000 psig, conditions which also promote some cracking reactions.

In the FMC base design, hydrotreating is indicated to be performed at a total pressure of 1710-1720 psia. Filtered oil from the filtration plant is pumped, along with hydrogen from a reforming plant and some recycled oil, through a gas-fired preheater into initial catalytic guard reactors. The guard reactors are intended to prevent plugging of the main hydrotreating reactors by providing for deposition of coke formed in the system on low surface-to-volume packing.

The hydrotreating reactors are indicated to be three-section, down-flow devices. The gas-oil mixture from the guard bed is introduced at the reactor head along with additional recycle hydrogen. Recycled oil and hydrogen at low temperature (100-200°F) are introduced between the catalyst sections in the reactor to absorb some of the exothermic heat of reaction.

The hydrotreated effluent is cooled and flows into a high-pressure flash drum, where oil-water-gas separation is effected. About 60 percent of the gas which separates is recycled by compression to the hydrotreaters. The remainder is indicated to be directed to the hydrogen plant.

A little less than half of the oil which separates is recycled to the hydrotreaters. The remainder, taken as product, is depressured into a receiving tank. From the tank it is pumped into a stripping tower, where clean product gas is used to strip hydrogen sulfide and ammonia.

Clean product gas is used also to strip ammonia and H_2S from the water which separates from hydrotreater effluent. Stripped water is indicated to be recycled to the last pyrolysis stage. The gas effluents from the strippers are indicated to be directed to gas clean-up.

The only major effluents to atmosphere from this section are the combustion gases from the hydrotreater preheater. About 4.5 tph of product gas is consumed, along with about 84 tph of combustion air. The products of combustion should be dischargeable directly without further treatment.

The process basis includes a large cooling requirement for hydrotreating effluent, even though preheating is supplied to hydrotreating feed. The developers (14) have indicated that heat integration should be possible in a commercial installation to some degree. The concern involves possible degradation of raw oil feed in a heating system which is not precisely controlled. We have assumed that 380,000 lb/hr of 600 psia steam will be generated in this cooler.

The process design basis does not provide for catalyst replacement in this section. Nor are facilities included for presulfiding catalyst, if this be required, or for regenerating catalyst. A major unresolved process question relates to the catalyst life that may be expected in commercial operation. Pilot plant results show that activity drops after 300-500 lb oil/lb catalyst, but pilot-plant conditions are considered more rigorous than should be the steady-state condition of the commercial unit.

Since high-temperatures are required generally for the regeneration of the cobalt molybdenum or nickel/tungsten sulfide catalysts used, we have assumed that regeneration, if it is practiced, will occur off-site. Moreover, we have assumed that the hydrotreaters will be designed to run continuously between maintenance shut-downs. It is not clear, however, whether two vessels provided are required to treat the total stream, or whether one represents stand-by capacity. Presumably some standby capacity will be required to permit catalyst changeout in the event of sudden activity loss or development of high pressure drop.

Provisions for depressuring and inerting the hydrotreater preliminary to catalyst removal should not result in emissions to atmosphere, since gaseous effluents may be recycled to the hydrogen plant gas treatment section, or to the main gas-treating section. Ammonium sulfide, which is produced in the hydrotreater, and which is stable at reaction conditions, decomposes at low temperatures and pressure to release additional ammonia and H_2S into the inerting medium. Metal carbonyls may also be present, and special precautions may be required if these are found in significant concentration.

Gaseous effluent which results from inerting the system after catalyst replacement may require treatment to remove particulates. Catalyst presulfiding may also produce gas which must be treated, although it is not yet certain whether and to what degree presulfiding improves catalytic activity. In general, the same procedures used to replace catalyst in the hydrotreater may also be applied to changeout of the packing or catalyst in the guard reactors. Presumably, more than one of these reactors will be provided for each hydrotreater to permit coke removal and bed replacement on the run.

3. EFFLUENTS TO AIR - AUXILIARY FACILITIES

We have elected in this study to treat the main conversion streams separately from all other facilities, which are thereby defined as auxiliary facilities. The functions of these auxiliary facilities are nonetheless required by the process, and, for economic and/or ecologic reasons, would be constructed along with the conversion system in an integrated plant. These effluent streams are also shown in Figure 2, and streams are identified in Table 1.

3.1 Oxygen Plant

The oxygen plant provides a total of 3760 tons per day of oxygen to the last pyrolysis stage. The only effluents to the air from this facility should be the other components of air, principally nitrogen About 340 MM scfd of nitrogen will be separated. Some of this nitrogen may be used to advantage in the plant to inert vessels or conveyances, to serve as transport medium for combustible powders or dusts, as an inert stripping agent in regeneration or distillation, or to dilute other effluent gas streams. Nitrogen is also indicated to be used to pressurize the rotary pressure raw-oil filters.

About 440 MM scfd of air is taken into the oxygen facility. Placement of the oxygen facility will depend in part on the desire to maintain the quality of the air drawn into the system and, especially, to minimize interference from plant effluents.

3.2 Acid Gas Removal

The acid gas removal process to be used in this facility has not been specified by FMC. Sulfinol and hot carbonate have been tentatively considered (30).

The primary feed to this unit would be the product gas stream separated from the product recovery system (513 tph). Contaminated product gas used for stripping the water and oil effluents from hydrotreating (107 tph) may also be returned to this unit, although this stream contains ammonia, and it may be preferable to treat it separately.

The particular choice of acid gas removal process may depend on the nature and quantity of "trace" contaminants present in the gas to be treated. Hence COS, if it is present, is hydrolyzed in the Benfield hot carbonate system to H_2S (31). Similarly, mercaptans disulfides, and thiophenes are indicated to be largely removed in commercial installations. FMC has not reported on the quantity and nature of the sulfurous contaminants in raw gas. COS has been found in some streams (14), and additional work is planned to quantify this and other trace constituents in COED streams.

Choice of process will, of course, also depend on installation and operating costs, as well as on the ease of integration into the total facility (32, 33). Most purveyors of proprietary processes can tailor their designs to accommodate particular conditions and requirements. Moreover, leading processes are being continuously refined and developed.

We have elected to use the "Benfield" hot potassium carbonate system in our basis. This method for removing CO_2 and H_2S from process gas streams was studied at the Bureau of Mines (34, 35, 36). In the Benfield system, gas absorption takes place in a concentrated aqueous solution of potassium carbonate which is maintained at above the atmospheric boiling point of the solution (225-240°F) in a pressurized absorber. The high solution temperature permits high concentrations of carbonate to exist without incurring precipitation of bicarbonate.

Partial regeneration of the rich carbonate solution is effected by flashing as the solution is depressured into the regenerators. Low-pressure steam is admitted to the regenerator and/or to the reboiler to supply the heat requirement. Regenerated solution is recirculated to the absorbers by solution pumps. Stripped acid gas flows to the sulfur recovery plant after condensation of excess water. Depressurization of the rich solution from the absorber through hydraulic turbines may recover some of the power required to circulate solution.

Raw product gas from the product recovery section must be compressed for effective scrubbing. The actual pressure level that will be employed will be a trade-off between compression costs and the utilities consumptions required otherwise. Based on the concentration of acid gases present in raw gas, a total scrubbing pressure between 100 and 200 psia is indicated, whether an amine or hot carbonate system is employed. We have estimated that the compressor driver will require the equivalent of 500,000 lb/hr. of high-pressure steam to handle the primary raw gas stream. Some 1,400,000 gph of solution must be circulated, requiring the equivalent of 5700 KW. Some 450 MM Btu/hr is required for regeneration, supplied as steam, and about this same cooling duty will be required. Additionally, some 100,000 lb/hr of high-pressure steam, 1200 KW and 95 MM Btu/hr as low-pressure steam and as cooling water will be required to treat the stripping gas stream.

Clean gas may be directed to the various fired heaters throughout the plant, and to the utility boiler (see below). Product gas loss into the regenerator off-gas stream can be held to less than 0.1 percent in proprietary configurations of the process. Moreover, it is possible to selectively remove H_2S, if this is required to produce a suitable feed for a Claus sulfur plant. There should be no discharge to the atmosphere from the acid gas removal section.

3.3 Hydrogen Plant

The COED process gas product is indicated to be the source of hydrogen for the hydrotreating of raw COED oil. Steam reforming, cryogenic separation, and partial oxidation were investigated by Chemical Construction Company (5) as means for recovering the required hydrogen from process gas. The type of hydrogen plant that may ultimately be used will be a function of the location of the plant (or of the coal type being processed) and of the product sales slate, as well as of the size of the installation. For our design, we have assumed the steam reforming case as outlined by FMC.

COED process gas at 15 psia is compressed to 410 psia and passed through a sulfinol system to remove CO_2 and H_2S. Regenerated acid gases are directed to the sulfur recovery plant. The cleaned process gas containing about 1 ppm H_2S is divided into a fuel gas stream and a process feed gas stream. The process feed gas is passed over a zinc oxide sulfur guard bed to remove sulfur traces, and is then heated by combustion of the fuel gas and hydrogenated with recycle product hydrogen to remove unsaturates. Steam is injected and reforming and shifting occur catalytically according to:

$$CH_4 + H_2O \longrightarrow CO + 2H_2 \quad \text{(reforming)}$$
$$CO + H_2O \longrightarrow CO_2 + H_2 \quad \text{(CO shift)}$$

CO_2 formed in the reactions is removed in a second scrubber-absorber and the process gas is finally methanated catalytically to convert residual CO to methane according to $3H_2 + CO \longrightarrow CH_4 + H_2O$. Resulting product gas is available at 200 psig.

The bleed gas from the hydrotreating plant, containing about 2 percent H_2S and about 0.1 percent ammonia, is indicated to be returned to the hydrogen plant for reprocessing. It may be preferable to first scrub this stream with water separately to remove the ammonia trace. About 3.5 tph of H_2S must also be removed from this stream, and the H_2S residual, after water scrubbing, would be removed in an acid gas scrubber and directed to the sulfur recovery plant.

About 9.4 tph of hydrogen is indicated to be consumed in hydrotreating 185 tph of raw oil (about 3000 ft^3/bbl). It is of course not required that initial acid gas removal be included in the hydrogen plant if acid gas removal is otherwise provided for the total product gas stream. Moreover, gas from the cleaning operation would be available at pressure, so that compression is required only from that pressure level. About a third of the hydrogen requirement can be generated from excess CO and hydrocarbons present in the hydrotreating bleed stream. About 25 tph of clean product gas would be required additionally to be fed to the unit, and about 43 tph of water would be consumed in the reformer.

If a hydrogen plant design as described is employed, it should be possible to recover energy from the expansion of the hydrotreating bleed gas through use of turboexpanders or equivalent facilities to offset the energy required for recompression to the level required in the hydrogen plant.

The major gaseous effluents from the hydrogen plant will be the products of combustion from the fired heaters and the CO_2 stream removed from the processed gas after reforming. Since clean product gas is consumed in the heaters, the products of combustion should be dischargeable directly. Some 23 tph of gas is fired.

About 60 tph of CO_2 will be removed from the process gas, and this too may be discharged, although there may be incentive to recover some or all of this stream for sale, since its purity should be high.

3.4 Sulfur Plant

The type of sulfur plant that will be used has not been specified by FMC. The combined acid-gas streams resulting from treatment of raw product gas (pyrolysis gas) and hydrotreating bleed gas would appear to yield an H_2S concentration of about 7 percent, based on gas analyses presented in the FMC design. Additional concentrated H_2S streams may result from treatment of sour water and stripping gas. FMC has indicated that high-sulfur Illinois coals will yield H_2S levels in the range of 10-20 percent (30).

We have assumed that acid gas will be sufficiently high in H_2S content to permit use of a Claus recovery system. We note that, depending on the acid gas removal process employed, H_2S may be preferentially absorbed to increase its concentration in off-gas fed to the sulfur plant. Claus units are operated commercially with entering H_2S concentrations as low as 6 percent. But these systems generally employ oxygen, so that some of the cost advantage relative to a process like Stretford, which does effectively treat low concentrations, may dissipate.

Tail gas from the Claus unit must be desulfurized, however. Several processes have been developed for this purpose. FMC indicates that the Beavon or Shell Claus Off-Gas Treating (SCOT) process may be employed (30). It may also be feasible to employ one of the flue-gas desulfurization variants using limestone to scrub tail gas (37-40), or processes such as the Wellman-Lord SO_2 Recovery Process (41) or the IFP Secondary Recovery Process (42) may be applied.

Most proprietary tail-gas treatment processes operate to convert SO_2 to H_2S, which may then be selectively removed. The Beavon system catalytically hydrogenates the SO_2 over cobalt-molybdate. The catalyst is also effective for reacting CO, which may be present, with water to form hydrogen, and for the reaction of COS and CS_2 with water to form H_2S.

The hydrogenated stream is cooled to condense water, and the H_2S stream is fed into a Stretford unit to recover sulfur in elemental form. Treated tail gas may contain less than 200 ppm sulfur, with almost all of this being carbonyl sulfide. Condensate may be stripped of H_2S and directed to boiler feed water treatment.

About 500 tpd of elemental sulfur will be separated at the sulfur plant, depending on the sulfur content of the feed coal and on the processing employed. Total sulfur emission to the atmosphere may be held to less than 200 lbs/hr., and the treated tail gas may be directed to a boiler stack for disposal. The small air stream used to regenerate the Stretford solution in the tail gas treatment plant may also be so directed.

3.5 Utilities

3.5.1 Power and Steam Generation

The choice of fuel for the generation of the auxiliary electric power and steam required by coal gasification plants markedly affects the overall process thermal efficiency. It is generally least efficient to burn the clean product gas for this purpose. On the other hand, investment in power-plant facilities, including those required to handle the fuel and to treat the flue gas, is generally least when product gas is so used.

COED conversion generates a carbon-containing char equivalent to some 50-60 weight percent of the coal fed to pyrolysis. Since this is considered a fuel product, it would appear that it should be so used in the plant proper. However, it suffers as an acceptable fuel in this case to about the same extent as does the feed coal, in that its sulfur content is observed to be about the same as that of feed coal. Table 2 lists the analyses of chars obtained by FMC from a low-sulfur western coal and from a high-sulfur Illinois coal.

For a high-sulfur coal feed such as an Illinois-No. 6 seam, combustion of the char produced will generate SO_2 flue-gas levels above permissible discharge limitations, such that some form of flue-gas treatment must be employed.

The char obtained from COED is also a more refractory material than feed coal. Char from Utah A-seam coal (a low-sulfur Western coal which can be directly combusted without recourse to sulfur controls), when pulverized, has much the same combustion characteristics as some Pennsylvania anthracites, and has been satisfactorily combusted in an anthracite boiler (43). Its low sulfur content and lower grinding power requirements would in fact command a premium over some anthracite coals.

Experience with the large-scale combustion of chars is otherwise limited. The Bureau of Mines has reported on one investigation (44) utilizing a specially-constructed dry bottom unit designed to simulate the performance of an industrial steam-generating furnace. In general, it was found that volatile matter content in excess of 20 percent was necessary for combustion of chars in this apparatus in the absence of a more volatile supplemental fuel (natural gas was used as supplemental fuel). Carbon combustion efficiency was likewise found to be a function of the volatile matter content of char, ranging from 94 to 99 percent for volatile contents from 5 to 15 percent. More supplemental fuel was required for the least volatile chars to maintain flame stability.

TABLE 2

PROPERTIES OF CHAR PRODUCT (30)

	Utah	Ill. No. 6
Proximate Analysis, wt %, dry		
Volatile Matter	6.1	2.7
Fixed Carbon	80.2	77.0
Ash	13.7	20.3
Ultimate Analysis, wt %, dry		
Carbon	81.5	73.4
Hydrogen	1.3	0.8
Nitrogen	1.5	1.0
Sulfur	0.5	3.4
Oxygen	1.5	1.0
Ash	13.7	20.3
Chlorine	0.006	0.1
Iron*	0.28	
Higher Heating Value, Btu/lb. dry	12310	11040

* Included in "ash" above

Hence, it may be assumed that combustion of COED chars will be possible in conventional fireboxes if clean product gas is used as supplemental fuel. This alternative might be preferred then on the basis of carrying the least developmental debits, and because it may be possible to adjust SO_2 concentration in flue gas such that subsequent flue gas treatment may be avoided. It has the disadvantage of adversely affecting overall thermal efficiency.

A further study by the Bureau of Mines (45) in the same dry-bottom unit has shown that a COED char derived from Illinois No. 6-seam coal and containing 5 percent volatile matter could be successfully fired without supplementary fuel if the char-primary combustion air mixture were preheated to 450-500° F. It was estimated that the heat required to raise the mixture from 100° to 450°F was equivalent to about 2 percent of the heating value of the char, whereas natural gas equivalent to 15 percent of the total thermal input would be required to stabilize combustion in the absence of preheat.

Ideally, the sensible heat of hot char discharged at the fluidized-bed char cooler-steam generator would be conserved in any subsequent char treatment process. Both the anthracite boiler test referred to above and the Bureau of Mines work employed chars which had been ground to pass 90-95 percent through 200 mesh screens. In the Bureau's work, at least, some slight decrease of combustion efficiency and increase in supplementary fuel requirement was noted when the degree of pulverization was decreased from 95 to 90 percent through 200 mesh.

Conventional grinding equipment installed on coal-fired boilers is generally designed to handle coal at less than 300°F. However, commercial equipment is available which can operate at higher temperatures, up to about 500°F. A system might be devised to heat air (by exchange with 800°F hot product char) that would be used as primary combustion air in the boiler. Or equipment may be designed to generate steam if char must be cooled for grinding.

The particular COED char employed in the Bureau's tests, although derived from an Illinois No. 6 coal, had a volatile matter content of 5.0 percent, or about double the level reported most recently by FMC (see Table 2) for char product from Illinois coal. Additional supplementary fuel or higher preheat temperatures may be required as the volatile content of char decreases.

Hence it is considered that equipment can be developed or modified to combust COED char at carbon combustion efficiencies above 95 percent, and that a large fraction of the sensible heat of product char may be conserved if the combustion is performed onsite, or at the point of production. There would, of course, be energy debits associated with the treatment of stack gases or with the use of specialized combustion systems, as by combustion in the presence of limestone in fluidized beds (48), to control sulfur emissions. And combustion of all of the char might support a 1200 MW generating station in this case.

A viable alternative to char combustion is its gasification, in which case sulfur recovery and water treatment are simplified and all fuel products produced are clean. Of course, facilities required to gasify char will add considerably to facilities provided otherwise, including acid gas removal systems, sulfur plant, and possible gas conversion facilities and/or oxygen plant.

A variety of proprietary processes have been considered for the gasification of COED chars, including the Koppers-Totzek process (46) and the COGAS system (47) under development by FMC. The Koppers process employs oxygen in a low-pressure, high-temperature gasifier to avoid nitrogen dilution of product gas, whereas COGAS maintains flue gases from the air combustion of char (which supplies heat to the gasifier) separate from the gasifier output. A number of other processes (49,50) may also be applicable.

All such processes, including processes used to upgrade gasifier output, involve thermal debits, however, such that there would be incentive on this basis only to produce a clean fuel at the lowest thermal cost. In a real situation, the product must be tailored to the consumer, so that economic and ecologic credits may outweigh the thermal losses.

We have in this study considered that dirty fuels would not be combusted in the plant, so that clean product gas would be used also for the generation of steam and power requirements. However, the total utility balances require some additional fuel source. Of the 513 tph of contaminated product gas issuing from the product recovery system, there is net 171 tph of dry gas available from the acid-gas removal system. Some 25 tph is required as feed to the hydrogen plant, so that the net available gas for fuel is 146 tph. The gas is estimated to have a higher heating value of 505 Btu per scf, so that the total available fuel gas equivalent is about 4180 MM Btu per hour.

Net steam requirements for the facility total 783,000 lb/hr, equivalent to a 1130 MM Btu/hr fuel requirement (See Table 10). Net electrical power requirements total 93,200 KW, equivalent to 902 MM Btu/hr of additional fuel. The plant otherwise fires fuel equivalent to 2842 MM Btu/hr in process heaters. Hence the total requirement, 4874 MM Btu per hour, cannot be supplied by the product gas stream alone. The shortfall, equivalent to 694 MM Btu/hr, would presumably come from char.

We have considered that the 2032 MM Btu/hr fuel equivalent required at the power plant could be supplied by the combinative firing of product char and product gas in suitably designed boilers per the Bureau of Mines work cited above. The fuel requirement is such that if all of the char required to supply the fuel shortfall, about 30 tph, is fired in the power plant along with about 47 tph of product gas,

the sulfur emission would be such that flue-gas treatment would still be required. About 2.1 tph of SO_2 would be emitted, equivalent to about 2.0 lb/MM Btu, or above the level permitted by current standards for solid fuels.

Flue-gas treatment might be avoided if char were combusted with product gas throughout the plant. This would require additional investment in char handling and grinding equipment, as well as particulate control on all fired heaters and ash handling and disposal facilities, and may be less attractive than installation of flue-gas treating facilities on the main boiler. A variety of flue gas treatment processes for particulate and SO_x control are under development (67,68), and significant progress in this area may be expected by the time a commercial plant is constructed.

We note also that the coal fines estimated to be produced in the coal grinding operation could supply the fuel shortfall. This alternative may be attractive in a commercial facility because there would be no additional grinding debit and because the fines production might be entirely consumed. However, such coal fines may command a higher premium as a saleable fuel than char, and it may be preferred to charge the coal fines to char gasification, depending on the system used for that purpose.

We have assumed for the purpose of thermal efficiency calculations that char will be combusted in the plant to make-up the fuel shortfall, and have not debited the process for flue-gas treatment. We recognize that char treatment (gasification) is practically required in a commercial design, and the effect of using char-derived gas for fuel is discussed in Section 5.

3.5.2 Cooling Water

A total of 200,000 gpm of cooling water is indicated to be required for operating the FMC design. Because most of this requirement is used for thermal exchange against relatively low-pressure streams, the circuit should be relatively free from process contamination leakage.

A design wet bulb temperature of 77°F and an approach to the wet bulb temperature of 8°F was assumed, with a circulating water temperature rise of 30°F. 9,000 gpm is required as cooling tower make-up, equivalent to 4.5 percent of circulation. Some 3,000,000 pounds per hour of water is evaporated at the cooling tower, 600 gpm is lost as drift, and 2400 gpm is withdrawn as blowdown, and is directed to the water treatment facility.

We have not included the cooling requirement to condense water from the coal grinding effluent gas stream. If water availability is constrained, this may be attractive. The plant water balance is shown in Table 11.

It is probable that environmental considerations and the costs of water reclamation will operate to restrict industrial water consumption in most domestic locations. Hence a commercial design might maximize use of air-cooled heat exchangers, reserving the use of cold water only for "trim-cooling" or low-level heat transfer applications. The overall economic balance will consider added investments in heat-exchange and electrical hardware associated with air-fin usage, as well as investment in incremental electrical generation capacity. Running costs for the generation of power and for equipment operation would be balanced against the net reduction in water treatment and pumping costs, as well as the net reduction in water loss.

On the basis that half of the requirement may be displaced with forced draft air-cooled heat exchangers, the incremental electrical power requirement is estimated to amount to 26,000 KW. Added cooling water requirement associated with the incremental power generation would bring the net total cooling water requirement to an estimated 100,000 gpm, so that water loss by evaporation might be reduced to about 3025 gpm at the cooling towers. Drift loss would amount to 300 gpm on this basis. Blow-down, or draw-off from the system, might be held to 1200 gpm. There would be a reduction in the power requirement for pumping cooling water. On the other hand, direct discharge of heat to the air environment in certain locations may be less desirable than the humidification associated with cooling towers.

The physical environmental situation at a particular site, including water availability, climatic conditions, and available area, will set limits on the designer's options for heat rejection. Other means, such as cooling ponds, may be practicable. In very special situations, it may prove economic to recover some of the low-level heat, as by circulation in central heating systems to nearby communities or in trade-off situations with irrigation water supplies, where hot water may be used to extend growing seasons. In all situations, the sociological impact of the use of the environment will be an over-riding factor.

3.5.3 Water Treatment

Analyses of the aqueous condensates produced in the pyrolysis and hydrotreating plants have not been specified in the FMC design. Some characteristics of these streams have been recently reported by FMC (See Table 3). FMC has also indicated that these streams would be preferentially recycled to the last, or hottest pyrolyzer, or to char gasification if it be included, after minimal processing to strip ammonia and hydrogen sulfide.

Table 3

Properties of Process Liquors (30)

Coal	Illinois No. 6
	First Stage Pyrolysis Liquor weight percent
Carbon	--
Nitrogen	0.05
Sulfur	0.07
Phenol	0.00
Entrained Oil	--
Suspended Solids	0.49
pH	3.6
	Second Stage Pyrolysis Liquor
Carbon	--
Nitrogen	0.93
Sulfur	0.18
Phenol	0.38
Entrained Oil	0.0-0.5
Suspended Solids	1.09
pH	8.8
	Hydrotreating Liquor
Carbon	0.8
Nitrogen	5.0
Sulfur	8.7
pH	9.3

Recycle to a high-temperature char gasification system should present no difficulty (46). However, the long-term recycle to pyrolysis requires additional study, since temperatures are rather low and there is no basis on which to estimate the degree of "by-pass" through the fluidized bed system. Demonstration of such long-term recycle, however, would considerably reduce investment in treatment facilities. The question may be largely academic, however, because it would appear that a large-scale installation, unless it were arranged to combust char onsite or in an adjacent facility, would include some form of high-temperature char gasification. We have assumed that pyrolysis liquor may be recycled in our design.

Facilities required to treat water, including raw water, boiler feed water, and aqueous effluents, will include separate collection facilities:

- Effluent or chemical sewer
- Oily water sewer
- Oily storm sewer
- Clean storm sewer
- Cooling tower blowdown
- Boiler blowdown
- Sanitary waste

Retention ponds for run-offs and for flow equalization within the system will be required. Run-off from the paved process area could easily exceed 15,000 gpm during rainstorms. Run-off from the unpaved process and storage areas could exceed 80,000 gpm in a maximum one-hour period.

Pretreatment facilities will include sour water stripping for chemical effluents and Imhoff tanks or septic tanks and drainage fields for sanitary waste.

Gravity settling facilities for oily wastes will include API separators, skim ponds, or parallel plate separators.

Secondary treatment for oily and chemical wastes will include dissolved air flotation units, granular-media filtration, or chemical flocculation units.

Oxygen demand reduction may be accomplished in activated sludge units, trickling filters, natural or aerated lagoons, or by activated carbon treatment.

Boiler feedwater treatment will in general involve use of ion-exchange resins. Reverse osmosis, electrodialysis, and ozonation may find special application.

We consider that the COED plant may be able to take advantage of the properties of char and of attractive incremental costs for oxygen to assist its waste water treatment. Hence, the char produced by the process may have some of the attributes of activated carbon (63), which has been shown to be effective in the removal of a wide variety of the water contaminants expected (64).

Similarly, oxidation of contaminants in water using oxygen, and especially ozone, is normally reserved for polishing drinking water supplies because of high costs. Direct oxidation, however, is very effective in reducing phenol, cyanide and thiocyanate levels in waste water (65), and has particular advantage in that solids concentrations are not thereby increased.

Evaporation will of course occur throughout this system, and the concern of the designers will be to limit the co-evolution of noxious or undesirable components which may be present. We note that it may be necessary to cover portions of the water-treatment facility and/or provide forced draft over some units to avoid undue discharge of hydrocarbons into the atmosphere. In the latter case, as with direct oxidation or ozonation, sweep gases would be ducted to an incinerator or boiler, and provisions for minimizing explosive hazard would be required.

3.5.4 Miscellaneous Facilities

Provisions for start-up of the COED facility may generate short-term effluents to the atmosphere. Reverse flow from the gas product delivery line may be practicable for fuel supply, or a pressurized gas storage facility might be provided on site.

Planned noise reduction, especially in coal handling, grinding, and charging operations, venting, and in the operation of large compressors and pumps, will be a requirement.

Operation of a blow down system and flare stack to which accidental or emergency process releases may be directed will normally produce a small emission to the atmosphere. We note that future effluent limitations may restrict all emergency hydrocarbon emissions, in which case the emergency flare system must be sized to handle the entire gas output.

4. LIQUID AND SOLID EFFLUENTS

Solid and liquid effluents based on our design are also shown in Figure 2 and Table 1.

4.1 Coal Preparation

On-site coal storage will require that facilities for containing storm run-off be provided. Hence, run-off from the 40-50 acre area required to hold a thirty-day supply of feed coal could easily amount to 10,000 gpm during a major precipitation event common to almost all sections of the United States. Such run-off may be expected to contain acidic particulate matter from most contemplated feed coals.

It is assumed minimally that effluent limitation guidelines published by EPA for the coal mining industry under the Refuse Act Permit Program (58) will apply to such coal storage facility. The application of 'best practicable control technology' would require installation of impounding and settling facilities to be of sufficient size to handle run-off resulting from a once-in-ten-years' storm, and the operator would provide suitable recording analytical equipment, including a recording rain gauge, to guarantee compliance with concentration schedules for discharges into waterways.

Since permissible concentration schedules are such that impounded water will, after treatment, be of sufficient purity to be admitted to the plant's water system, it will be advantageous to plan for such use in the initial design. Similarly, run-off from the pyrolysis complex otherwise will have to be contained. More than one set of water treatment facilities will be required to handle the various water streams coming from and going to the plant. Depending on the severity of contamination that may be expected from the various processing areas, storm run-off from such areas would be directed to segregated holding facilities consistent with the expected water quality (See Section 3.5.3). It may be necessary to provide an impermeable subsurface barrier under certain portions of the facility, as the coal storage area, to prevent contamination of ground water.

Although not necessarily considered a part of the conversion facility, the coal mining operation, if it be located adjacent to the gasification complex, would probably share treatment facilities provided for the plant proper. Hence, typical acid mine drainage, of perhaps 300-400 gpm (59), might be treated continuously by accepted techniques (60,61), to produce water suitable for discharge or for plant use. Except for a separate initial holding pond and small lime addition facility, all other components of the treatment facility would amount to incremental increases on facilities which must be provided the parent plant.

If coal laundering is practiced, facilities for retention and disposition of liquid and solid effluents become more complex. Combinations of screening and thickening devices will generate streams of varying solids content, some of which would be considered refuse, and have to be returned to the mine or be buried otherwise. A properly designed system would minimize make-up water requirement by internal treatment and recirculation of wash water. A facility to launder feed coal for this design might circulate 3700 gpm of wash water, and discharge 3000 gpm along with thickened refuse. Such refuse would be impounded in clarifying basins, where evaporative loss would occur. Make-up requirements would then be held to such evaporative loss and to an estimated 500-700 gpm lost via the laundered coal product.

4.2 Coal Grinding

The fines generated in the coal grinding operation, amounting to 5 percent of the coal fed, issues as a separate fuel product. This material would be preferentially charged to the char gasification system, if this operation is included. Alternatively, it could be burned as fuel within the plant in combination with clean product gas.

4.3 Coal Drying and First-Stage Pyrolysis

The only major liquid effluents from this section are the aqueous streams purged from the scrubber circuit and resulting from the filtration of fines from the scrubber liquor. These combined streams, totalling some 93.5 tph, are indicated in the FMC basis to be directed to the last pyrolysis stage. An analysis of this stream has been recently reported (30), and is shown in Table 3. It will presumably require clarification and pH adjustment if it is not consumed directly in pyrolysis or char gasification. Alternatively, it could serve to scrub ammonia from hydrotreating bleed gas, and could then be directed to waste water treatment.

The fines filtered from the scrubbing circuit are contaminated with oil or tar. This stream, amounting to 22 tph, is indicated to be recycled to coal feed.

4.4 Stages 2,3,4 Pyrolysis

The only major solid effluent from this section is the char product, amounting to 521 tph. Depending on the system used to cool char, or to recover its sensible heat, additional streams may be generated.

4.5 Product Recovery

The major liquid effluent from this section is the waste liquor purge from the scrubbing circuit, amounting to 237 tph. The analysis of this stream has also been reported by FMC (see Table 3). It, too, is indicated to be preferentially returned to the last pyrolysis stage, or to char gasification if included.

We have considered that conditions in pyrolysis may not be sufficiently severe to consume the expected contaminants. However, the indicated utility requirement for treatment of the combined liquor streams, as by the application of a process such as the Chevron WWT process (62), is very large, so that we have not included such treatment in our design. FMC may develop more definitive information regarding such recycle (14). FMC has, however, indicated that a water treatment facility may be required to handle process upsets in any event (30).

4.6 COED Oil Filtration

The major effluent from the filtration plant is the filter cake, indicated to contain about 1.5 tph of filter aid, 5.8 tph of raw oil and 7.9 tph of char fines. This stream is now indicated to be recycled to the coal feed stream. A small amount of basecoat would also issue with this stream.

4.7 Hydrotreating

The major liquid effluent from this section is the waste water stream separated from hydrotreater effluent, amounting to 16.6 tph. The analysis of this stream is also shown in Table 3. FMC indicates that it would preferentially be added to pyrolysis liquor and recycled to the last pyrolyzer.

There is indicated to be a very small coke make in the guard reactors, which is added to the char product.

Disposition of spent hydrotreating catalyst and catalyst or packing from the guard reactors may require special procedures if metal carbonyls are present (see Section 4.8.9.)

4.8 Auxiliary Facilities

4.8.1 Oxygen Plant

About 17 gpm of water will be condensed from entering air at the oxygen facility. This water should be suitable for addition to the plant's boiler feedwater treatment system.

4.8.2 Acid Gas Removal

Condensate streams will be generated as circulated gas is cooled in the acid gas removal system. The disposition of these streams will depend on their composition, but they may in general be directed to the waste water treatment facility.

Facilities will be required to dispose of contaminated Benfield solution, if this system is used, but the vendor now indicates that this stream would normally be very small.

4.8.3 Hydrogen Plant

Assuming that steam reforming will be used to generate the hydrogen requirement, the only major liquid effluents will be the excess water condensed from circulated gas. Such condensates, depending on the point at which they are collected, may be quite pure and can be directed to the boiler feedwater treatment system. Others may require treatment to remove dissolved acid gases, and may then be directed to the waste water treatment facility.

Periodic replacement of catalysts employed in this section will generate solids streams whose disposition requires further study (see Section 4.8.9).

4.8.4 Sulfur Plant

Elemental sulfur make from the Claus unit is about 490 tpd. An additional 15-20 tpd can be recovered from the Beavon tail gas treatment facility. The precise quantity of sulfur produced is a distinct function of the sulfur concentration in feed coal.

The Stretford system used to recover elemental sulfur from hydrogenated tail gas in the Beavon process requires a small liquid bleed to prevent build-up of thiosulfate and sulfate salts which may impair recovery efficiency. The waste bleed is high in chemical oxygen demand, and is generally incinerated. The COD of this waste stream may be lowered by adding sodium as caustic. The caustic requirement in this case is quite low.

4.8.5 Power and Steam Generation

Because clean fuel is indicated to be consumed in the power plant, there should be no significant liquid or solid operating effluent streams (see Section 5). Blow down from steam boilers may be included as make-up to the cooling water system.

It is necessary to chemically clean the boiler and associated piping before it is placed in service, and at an average interval of 2-3 years thereafter (24). Both acidic and alkaline solutions are used in chemical cleaning. The acidic wastes would typically consist of solutions of hydroxyacetic and formic acids, or hydrochloric acid, at concentrations of less than 5 percent. The alkaline wastes would consist of dilute sodium phosphate solutions (less than 1 percent). A large amount of water would have to be used for flushing the system.

For a boiler of this size, the total amount of waste produced could amount to several hundred thousand gallons of acidic and alkaline solutions and up to a million gallons of flushing water. In power plants, these wastes may be routed to settling ponds (ash basins) where they may be diluted and neutralized prior to discharge. Alternatively, contract chemical cleaning specialists may provide off-site disposal services.

4.8.6 Cooling Water

A variety of chemical additives may be used to treat water circulated in the cooling water system to control algae and corrosion. These will appear in tower draw-offs, along with matter originally present in make-up streams. Depending on the extent of facilities provided to treat waste water effluents, such draw-offs may be treated to precipitate or neutralize specific toxic elements such as chromium or zinc before being directed to further treatment.

Heat exchangers in cooling service may require periodic chemical cleaning, and facilities for disposition of chemical cleaning wastes will be required (see Section 4.8.5).

4.8.7 Miscellaneous Facilities

A variety of materials may be required to treat waste water effluents, including antifoam, phosphoric and sulfuric acids, and char or activated carbon. In addition, water treatment may require the use of lime-soda alums, ion exchange resins, caustic, ferrous ion, and chlorine, among other agents. Ultimately, these additives exit the system as concentrated sludges, contaminated solids, or in aqueous streams with high salt content. These effluents may be concentrated, dried, and/or incinerated. Ultimate disposition of the dry or concentrated residuals is uncertain, however, especially if heavy metals, leachable salts, or organic contaminants are present. Burial in sealed pits appears the only practicable method for disposal of materials which must be prevented from leaching into ground or surface water, although the logistics and economics of such techniques requires extensive further study.

4.8.9 Maintenance

Normal plant operations will require the periodic replacement or replenishment of catalysts and other chemical agents used to process gas and oil. Such maintenance will generate contaminated solid and liquid effluents, including shift catalyst, Benfield solution, activated carbon, zinc oxide, and caustic streams. In general, spent materials will be sulfidic. Metal value may justify specific reclamation, but again, it would appear that the ultimate disposition of such solid effluents is now uncertain. Incineration or thermal oxidation, as in a fluid bed incinerator, might be used to remove hydrocarbon and sulfur, but control of metallic particulates from such systems requires further study, as does the disposition of residues.

5. THERMAL EFFICIENCY

The determination of thermal efficiency is useful for providing a basis on which to compare like processes, or to gauge incentives for process improvements. Obviously, there are other equally important bases on which processes may be compared, including the economic efficiency, which may compare the total cost of a product or products, and the ecologic efficiency, which may compare the net irreducible pollution potential of a process. Obviously, all such comparisons must be performed on a common, well-defined basis, and all such comparisons are related to the technological state-of-an-art at a given point in time.

In the case of fossil fuel conversion processes, the thermal efficiency is calculated as the ratio of the heating value of product(s) to the heating value of the (coal) feed, assuming that coal is the sole source of raw material and energy. In the present design, the higher heating value for coal feed is reported as 12,420 Btu per pound. Product gas (clean, dry) has been estimated to have a heating value of 505 Btu per scf. Product char has been reported to have a heating value of 11,040 to 11,700 Btu per pound; we have used the higher value. Hydrotreated oil has been assigned a heating value of 19,100 Btu per pound based on an indicated 25° API gravity.

The fuel shortfall (see Section 3.5.1) may be supplied by burning product char. Assuming that 30 tph of product char is combusted in the facility along with the produced gas to supply the estimated total plant fuel requirement, a base thermal efficiency of 72.2 percent is indicated (see Table 4).

However, char is not a "clean" product in this case and should be discounted on some basis. If char were to be gasified in a Koppers-Totzek gasifier, the estimated gas yield is equivalent to 69 percent of the char heating value (46), so that the net overall efficiency is indicated to be reduced to 57.8 percent on this basis. The incentive, therefore, to develop an efficient char utilization process is very great.

In an integrated plant which includes char treatment, it should be possible to arrange the system such that coal is dried with combustion flue gases. There may also be economies possible in the treatment of water and of acid gases. These effects have been estimated to amount to the equivalent of about 600 MM Btu/hr, so that the net efficiency increases to 60.2 percent.

The thermal efficiency of the Koppers-Totzek char gasification process degrades significantly if product gas must be compressed for delivery. For example, if product gas could not be utilized at 15 psig, but had instead to be compressed to 150 psig, the Koppers-Totzek efficiency would drop to 61 percent, i.e., some twelve percent of the product gas equivalent would be consumed in the compression. The overall net efficiency would be about 56 percent in this case.

The developer prefers to consider that COED medium-Btu product gas would not be consumed in the plant, but that the excess over that required by the hydrogen plant would be sold (14). We have estimated a total of 48 tph for the hydrogen plant (feed plus fuel). Assuming that char is gasified in a Koppers-Totzek system, the char-derived low-Btu gas would be fired in the plant's heaters and in the power plant in this case. The revised product slate for this basis is shown in Table 4A. (Note that all values for gas tonnages given in the process description are referred to medium-Btu COED product gas). Any split between medium and low-Btu gas products will be essentially thermally equivalent if combustion efficiencies and delivery pressures are assumed identical. There will be slight thermal and economic debits associated with the firing and/or sale of low vs. medium-Btu gas.

The discrepancy in overall efficiency between the two product slates (Tables 4 and 4A) is due to the assignment of full heating value for char that is combusted to supply the fuel shortfall in the first case, without debiting the system for SO_2 removal from stack gases or for imperfect char utilization. In the second case, all char is converted to clean gas, and additional sulfur is recovered in elemental form, but at a 69 percent efficiency based on char feed to gasification.

The combustion of char in a system designed to limit SO_2 emissions may likewise be a good candidate for further development. Although a large number of flue-gas stack-treatment processes are undergoing active development, none has so far emerged as the industry standard. We have accordingly not attempted to apply the thermal debit to char combustion which may result from the application of such treatment. Again, it may be more to the point to combust char in a specialized facility, as in a fluidized bed in the presence of a limestone sulfur acceptor, to generate electricity in a combined-cycle operation (48).

The proprietary COGAS development (47) may also show an improved efficiency. This has been estimated to be 69 percent overall by the developers, compared with the 60 percent value estimated here for the coupling of Koppers-Totzek to COED.

Table 4

Thermal Efficiency

Medium-Btu Product Gas Fired in Plant Heaters and Boiler

	Quantity (tph)	Equivalent MM Btu/hr	Thermal Efficiency as Percent of Pyrolysis Feed	
Coal	1000	24,840	--	
Hydrotreated Oil	164.4	6,280	25.3	
* Product Char	491	11,505	46.3	
Sulfur	20.8	160	0.6	
	Base Efficiency			72.2
Char Gasification (69% thermal efficiency if low-Btu gas is made available at 15 psig)		-3565	-14.4	
	Net			57.8
Fuel Economies in Integrated Plant		+600	+2.4	
	Net			60.2

* Adjusted for char combustion to supply fuel requirements at 100 percent char utilization.

Table 4A

Thermal Efficiency

Char-Derived Gas Fired in Plant Heaters and Boiler

	Quantity (tph)	Equivalent MM Btu/Hr	Thermal Efficiency as Percent of Pyrolysis Feed
Coal Feed	1000	24,840	--
* COED Product Gas (505 Btu/SCF MW = 13.4)	123	3,520	14.2
Hydrotreated Oil	164.4	6,280	25.3
Gasifier Product Gas (318 Btu/SCF MW = 20.9)	364.3	4,200	16.9
Sulfur	40.0	310	1.2
			57.6
Fuel Economies in Integrated Plant		+600	+2.4
Net			60.0

* 48 tph product gas fed to hydrogen plant as feed and fuel.

6. SULFUR BALANCE

The sulfur balance for this design (See Table 5) suffers from the imprecision associated with the absence of consistent specifications for the sulfur content of feed coal, product char, oil, and liquor streams. Moreover, it would appear that the sulfur content of gas streams has not been completely specified, and may not be precisely representative for the assumed sulfur concentration in feed coal otherwise.

However, the sulfur balance, like the balances for other elements, requires only slight adjustment in the concentrations reported for large streams to close satisfactorily. On the other hand, the form in which sulfur appears in gas and liquor streams may have significant impact on the procedures and costs required to treat the streams. Only H_2S has been reported thus far, but a wide range of sulfur compounds would be expected to appear in pyrolysis gaseous and liquor effluents (62).

Sulfur content of Syncrude and sulfur emissions from the sulfur plant shown in Table 5 are fairly well-defined. The sulfur content of gas streams calculated from the FMC base design and the sulfur content of process liquor streams reported by FMC (30) are probably in need of adjustment to put them on a consistent basis.

Finally, a slight adjustment in the sulfur content of product char exerts a large influence on the overall balance because of the size of this stream. Whereas we have assumed a sulfur content for product char of 3.2 weight percent, our balance would indicate that char would have 3.7 weight percent sulfur content. The balance reported by FMC (30) would indicate a char sulfur content of about 3.4 percent on the same basis.

Table 5

Sulfur Balance (tph)

Illinois, No. 6 seam	Reported by FMC (30)	Estimated Per Design
Coal (Total Input)	41.0	41.0
Syncrude	0.2	0.2
Elemental Sulfur	22.4	20.8
SO_2 Emissions (1)	0.1	0.1
Char (to gasifier)	18.3	19.9

(1) From sulfur-recovery plants. Sulfur emission may be mostly in the form of carbonyl sulfide if Beavon tail-gas treatment is used. This balance assumes no sulfur emission in the purge gas stream from Stage 1 pyrolysis and recycle-to-extinction of aqueous process condensates. Additional sulfur emission approximately equal to the SO_2 value given above may be expected from the auxiliary gas cleaning facility of the char gasification plant.

7. TRACE ELEMENTS

Trace elements are usually defined as those elements present to the extent of 0.1% (1000 ppm) or less. Nearly all trace elements show an enrichment in coal ash relative to their crustal abundance (51). Manganese and volatile elements such as mercury are exceptions. This enrichment is attributed to concentration effects or exchange reactions during the formation of coals. Almost every element has thus been found in coals, but the variation in concentrations is quite broad (52).

The fate of trace elements present in the feed coal to conversion processes has so far received little attention. To the extent that such conversion processes approach conditions which obtain during combustion, it may be pertinent to apply results obtained in trace element studies of the combustion of coals (53-55). Even in such studies, however, the conditions of combustion have been noted to affect element dispositions. Coal handling and preparation methods can likewise influence results, so that generalizations may not be meaningful. Obviously, extrapolation to a particular conversion process or feed coal would be conjectural in large measure.

Although very large quantities of coal are consumed in combustion processes and the total quantities of trace materials, some of which are highly toxic, that may be released are likewise large, it has been only recently that concerted effort has been directed to the definition of the real problems. This effort, of course, has been associated with the promulgation of sanctions affecting permissible discharges to the atmosphere and waterways of the United States. Particular sanctions relating to toxic discharges are still in process of formulation (56). Research is required in many cases not only to set limits and goals, but also to develop analytical procedures that may be generally adapted. With fossil fuels, the general problem relates to the complexity of the chemical system, including the large number of components, the imprecision of available sensors or test methods, and the difficulties associated with representative sampling of very large streams. The detection and monitoring of many trace elements requires sophisticated procedures and equipment which cannot be practically applied commercially. In fact, the magnitude and nature of many industrial streams is such that direct quantification or measurement is impractical. The general nature of the pollution problem associated with COED Conversion has been described recently (30). At this point it is generally considered that COED will present no insurmountable control problems. On the other hand, additional research will be required to establish the degree of control which may be required.

Trace element concentrations in the gaseous and liquid streams that may be discharged to the environment from COED operations have not been reported by FMC. Of particular concern may be the purge gas

stream from the first stage of pyrolysis and the splits that may occur among the oil, aqueous condensates, and gas in the product recovery system. Toxic trace elements which may wash into the aqueous streams, for example, may require that such streams be specially treated if condensates are to be recycled to extinction to pyrolysis or to char gasification.

Each developer of a coal conversion process may ultimately be required to account for the disposition of elements present in feed whose toxicity or ultimate impact on the environment warrants control. He may moreover be required to guarantee the containment or neutralization of such materials in effluent streams, and this, in turn, may influence the adoption of particular processing alternatives. For COED this will require additional research firstly to define the levels of these elements through the process sequence for particular feed coals at preferred conversion conditions. A preliminary study of this type has been reported for a bench-scale coal gasification unit (57). Considerable amounts of many elements may be lost from ash during pyrolysis and gasification (See Table 6). Such loss may be appreciable, even though the processing temperatures employed may be relatively low. Information is required to detail the disposition of such losses, especially as they may appear in products and in process effluents. Moreover, the capacity of a large system to trap out various elements, as by chemical combination with materials of construction or through physical condensation, introduces another order of complexity, especially if process changes can result in sudden large emissions.

It would appear that COED does not introduce new control problems. Rather, since the pyrolysis train and water loops, including run-off, may be designed to be largely self-contained, emphasis of the controls development will be directed to the purge stream from the first pyrolysis stage, to the gas residual from acid gas treatment, and to the concentrated residuals from water treatment. Char gasification, if it is included, will present additional research needs (46). The enormous current government/industry effort to define and set effluent goals and to develop economical control procedures for coal-fired industrial operations will have a direct bearing on the extent of additional research that may be required, once stream compositions have been completely defined for COED conversion.

Table 6

Trace Element Concentration of Pittsburgh No. 8 Bituminous Coal At Various Stages of Gasification

Calculated on the Raw Coal Basis (From Ref. 57)

	Feed Coal	After Pretreat	After Hydro-Gasifier	After Electro Thermal Gasifier	% Overall Loss for Element
Max.Temp.of treat °C	-	430	650	1000	
Element:	ppm				
Hg	0.27	0.19	0.06	0.01	96
Se	1.7	1.0	0.65	0.44	74
As	9.6	7.5	5.1	3.4	65
Te	0.11	0.07	0.05	0.04	64
Pb	5.9	4.4	3.3	2.2	63
Cd	0.78	0.59	0.41	0.30	62
Sb	0.15	0.13	0.12	0.10	33
V	33	36	30	23	30
Ni	12	11	10	9.1	24
Be	0.92	1.0	0.94	0.75	18
Cr	15	17	16	15	0

8. PROCESS AND ENGINEERING ALTERNATIVES

Most of the process and engineering alternatives we have considered in connection with the particular design chosen as the basis for this report have already been presented or analyzed by FMC in the course of process development (1-6). The most far-reaching alternative involves the choice of fuel for the facility, and closely related is the treatment of char product. The net thermal efficiency of the process is largely a function of the alternatives chosen, clearly indicating the need for a definitive char treatment development.

Pyrolysis yields otherwise have been well-defined for a variety of coal feeds, so that these are not seen to be capable of significant change by process modification.

We consider the demonstration of long-term recycle of contaminated process liquors and condensates to pyrolysis to be equally important to the char treatment development. The necessity to process these streams with conventional sour-water stripping processes will add greatly to the utility requirement and investment in plant.

The oil absorption plan (5) for eliminating or reducing the oil filtration requirements could significantly affect investment, but is not likely to greatly influence the system otherwise.

The choice of system or systems for the removal of acid gases from the various product streams may have significant impact on utility requirements. As discussed in Section 3.2, available commercial systems are in continuous development, and there is every expectation that efficiencies will be improved.

Similarly, the choice of sulfur recovery method may be largely influenced by expected emission regulations, and could also bear heavily on investment and utility requirements.

The plant location will generally dictate the preferred method for heat rejection, which may significantly affect investment and make-up water requirements.

Table 7 lists some of the alternatives considered in connection with the base design.

Table 7

Process and Engineering Alternatives

Coal Drying

- Fuel-fired vs use of hot flue gases
- Venturi-scrubbing vs bag filters
- Catalytic CO oxidation vs stack dispersion

Pyrolysis

- Purge gas dispersion vs catalytic oxidation
- Hot-char water-cooling vs air-cooling

Product Recovery

- Aqueous scrubbing vs oil absorption
- Treatment of contaminated aqueous streams vs recycle to pyrolysis

Hydrotreating

- Preheat integration vs water-cooling
- Power generation on depressurization of hydrotreating bleed gas

Acid-Gas Removal

- Amine vs hot carbonate systems
- Separate treatment of main product gas stream vs combined streams

Hydrogen Plant

- Reforming vs cryogenic separation
- Amine vs hot carbonate CO_2 separation

Sulfur Plant

- Stretford vs modified Claus with Beavon tail-gas treatment

Utilities

- Alternative fuel choices for power and steam generation
- Waste-water treatment variations, including use of process char and oxygen or ozone
- Maximum air-fin usage vs cooling water for heat rejection

9. QUALIFICATIONS

This study is based on the process design (Tables 8 and 9 and Appendix Figures 1-5) supplied by FMC, the process developer, with modifications as discussed and shown in Figure 2 and Table 1. Costs or economics were not considered, except directionally.

Although mass balances presented in the flow sheets were found to be exact, it was not possible to achieve elemental balances overall. The flowsheets do not specify the elemental composition of coal, char, oil, or aqueous condensate streams. A number of varying analyses reported by FMC for Illinois No. 6-seam coal feeds, for the chars therefrom, and for the oils recovered were used in an unsuccessful attempt to achieve element balances.

Apparently, the analyses of gas streams shown reflect pilot-plant observations, but we were unable to key the analyses to particular pilot runs. Further, because the treatment of raw gas streams and the possible treatment of char is not specified, it was not possible to reconcile gas stream compositions otherwise. The relatively low total pressure and hydrogen requirement shown for hydrotreating filtered oil apparently reflect data obtained after December, 1972 (6).

We note that the average overall pilot-plant material balance (5,6) closed to within less than 5 percent, but the elemental balances were often poorer by factors of 5 or 6. In our report, we have not adjusted the compositions reported by FMC. Discretionary adjustments were necessary in some calculations.

Variations in feed coal and product compositions make it difficult to compare gasification processes. Significant variation is seen even for the "same" process on different coals. Similar variation will extend to the pollution potential of the process. Additional research and/or development will be required to define pollutant levels in particular streams with the precision required by today's standards, and so permit a more accurate assignment of energy requirements.

Table 8

FEED COAL AND PRODUCT CHAR ANALYSIS (15)

	FEED COAL	PRODUCT CHAR
Coal	Illinois	
Bituminous Rank (ASTM D386-38)	High-volatile C bituminous	
Seam	No. 6	
Mine	Peabody No. 10	
Type	Slope	
Town	Pawnee	
County	Christian	
Owner	Peabody Coal Co.	
Size, as rec'd., in.	1 1/4 x 1/4	
Moisture, as rec'd., wt. %	12	1.0
Proximate Analysis, wt. %, dry		
Volatile Matter	37.2	2.5
Fixed Carbon	51.1	76.3
Ash	11.6	21.3
Ultimate Analysis, wt. %, dry		
Carbon	66.9	73.8
Hydrogen	4.9	0.8
Nitrogen	1.1	1.0
Sulfur	4.1	3.2
Oxygen	11.7	0.0
Ash	11.3	21.2
Gross Heating value, Btu/lb.	12420	11700

Table 9

Typical Syncrude Properties* (30)

Coal Source	Illinois No. 6-seam
API, °@60°F	22
Pour Point, °F	0
Flash Point, PMCC, °F	60
Viscosity, cs. @ 100°F	5
Ultimate Analysis, wt. %	
C	87.1
H	10.9
N	0.3
O	1.6
S	0.1
Ash	<0.01
Moisture	0.1
ASTM Distillation	
IBP	190
10%	273
30%	390
50%	518
70%	600
90%	684
EP (95%)	746
Metals, ppm	<10
% Carbon Residue, 10% Bottoms	4.6
Hydrocarbon Type Analysis, Liquid Vol. %	
Paraffins	10.4
Olefins	0
Naphthenes	41.4
Aromatics	48.2

* Properties depend on severity of operation of hydrotreating unit.

Table 10

UTILITIES

	Cooling Water (GPM)	Power (KW)	Fuel Use (MM BTU/HR)	Steam (150 psi)	Steam (600 psi)
Coal Preparation	--	5,850	455.0	---	---
Stage 1 Pyrolysis	30,700	27,650	1144.0	5,000	---
Stages 2,3,4	--	40	298.0	-180,000	-265,000
Product Recovery	79,500	4,030	88.0	---	---
Oil Filtration	1,100	1,300	---	19,300	1,000
Hydrotreating	600	30,100	167.0	---	-380,000
Oxygen Plant	33,000	---	---	---	550,000
Acid Gas Removal	41,500	6,900	---	461,500	600,000
Hydrogen Plant	12,000	1,700	660.0	---	---
Sulfur Plant	1,000	1,300	30.0	-29,000	---
Power Plant	300	-93,170*	2032.0	-276,800	-506,000
Cooling Water	---	7,700	---	---	---
Water Supply and Treatment, Waste Disposal, and Misc.	300	4,400	---	---	---
TOTAL C.W	200,000				

* 2200 KW consumed internally.

Table 11

Plant Water Requirements

Users	GPM
Cooling Tower Makeup	9,000
Treated Boiler Feedwater (includes H_2 plant requirement)	1,550
Raw Process Water	-40 *
Potable Water	70
	10,580
Streams To Waste Treating	
Cooling Tower Blowdown	2,400
Boiler Blowdown	270
Oily Process Water	300
Sanitary	20
	2,990
Raw Water Makeup (Assumes 100% reuse)	7,590

* Net water make.

10. RESEARCH AND DEVELOPMENT NEEDS

FMC Corporation, in its development of Project COED), is well-advanced in terms of demonstrable process operability on a significant physical scale. Process yields have been well-defined for a variety of feed coals (1-6). And extrapolation to a commercial design is on a better basis in this case than for most other uncommercialized conversion processes under development.

Perhaps the most important research need is the development of an efficient char-utilization process. This aspect of COED development has received considerable attention already (4), and the COGAS developers (47) have two alternative gasification systems under study.

Another important research need relates to the treatment of contaminated process liquors and condensates. A large additional debit in thermal efficiency is seen if these materials cannot be recycled to extinction to pyrolysis, as is now assumed. This conclusion is based on analogies drawn with other processes, and may not prove to be the case here. More detailed analyses of the contaminants present in these streams than has already been reported (30) will permit a more accurate determination of overall treatment requirements. But the demonstration of long-term total recycle in the pilot-plant will serve better as a basis for design.

In this same connection, the "micro-structure" of gas, liquid, and solid streams requires further definition. The forms in which sulfur appears in these streams, and the toxic element contents may significantly affect expected dispositions and treatments. Future pilot-plant work should be directed to achievement of toxic trace element balances, especially for mercury, arsenic, cadmium, fluorine, and lead. Similarly, the concentrations of the various forms of sulfur should be established for all major gaseous streams, and for liquid streams which are, or may be, directed to treatment facilities.

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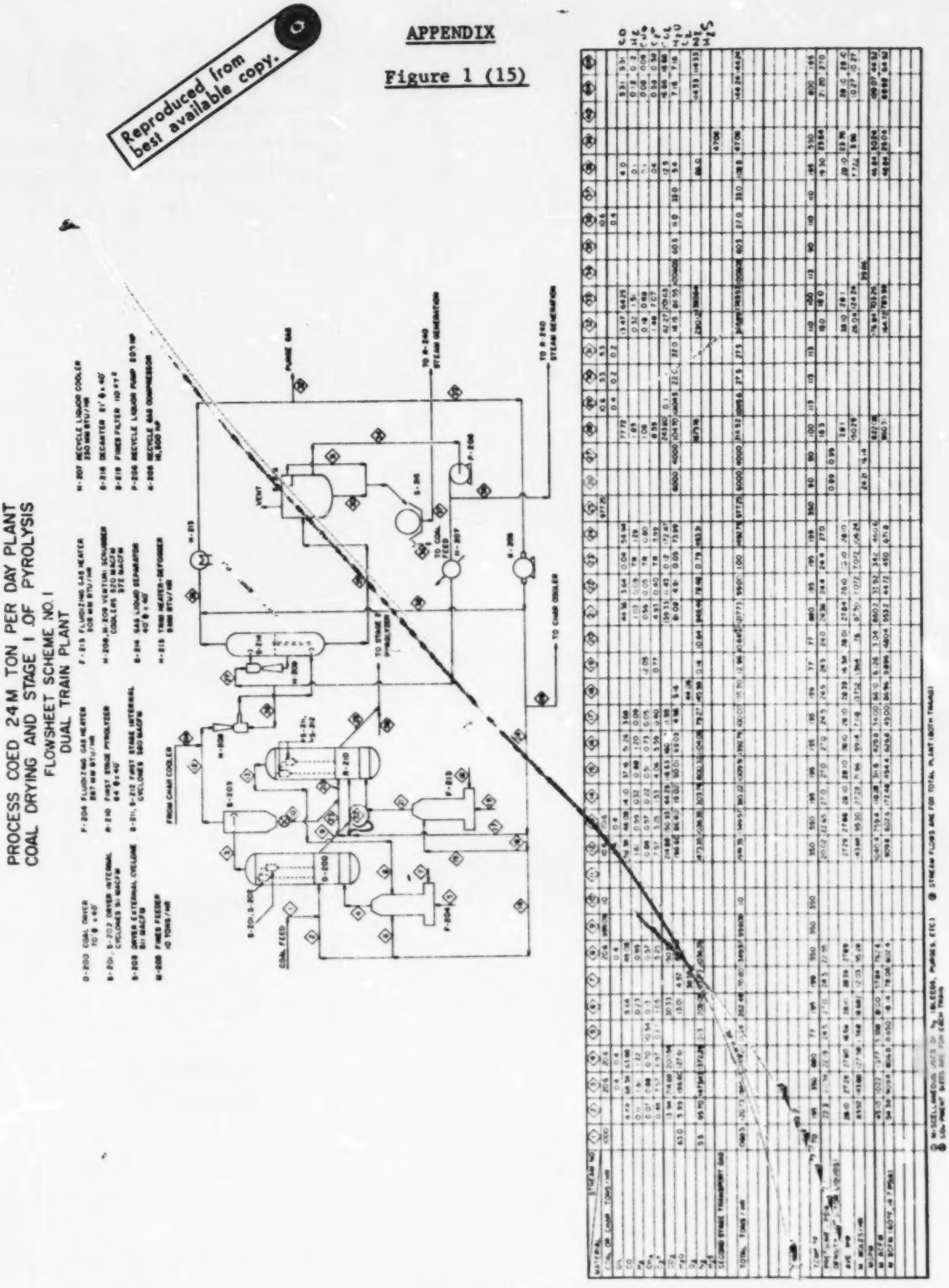
APPENDIX

Figure 1 (15)



PROCESS COED 24M TON PER DAY PLANT
STAGE 2, 3 AND 4 PYROLYSIS
FLOW SCHEME NO. 1
DUAL TRAIN PLANT

R-220
SECOND-STAGE PYROLYZER
S-221, S-222 SECOND-STAGE INTERNAL CYCLONES
S-223 SECOND-STAGE EXTERNAL CYCLONE

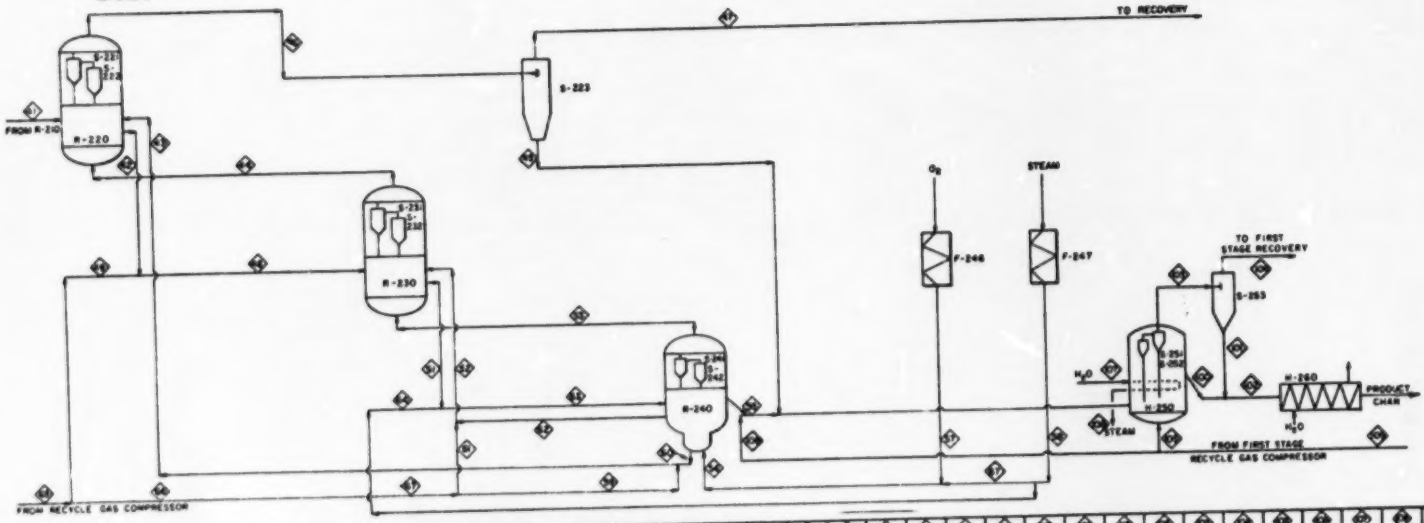
R-230
THIRD-STAGE PYROLYZER
S-231, S-232 THIRD-STAGE INTERNAL CYCLONES

R-240
FOURTH-STAGE PYROLYZER
S-241, S-242, FOURTH-STAGE INTERNAL CYCLONES

F-246
OXYGEN HEATER
F-247
STEAM SUPERHEATER

H-250 CHAR COOLER
STEAM HEATER
S-251, S-252 CHAR COOLER INTERNAL CYCLONES
S-253 CHAR COOLER EXTERNAL CYCLONE

H-260
CHAR COOLER



STREAM NO.	①	②	③	④	⑤	⑥	⑦	⑧	⑨	⑩	⑪	⑫	⑬	⑭	⑮	⑯	⑰	⑱	⑲	⑳	㉑	㉒	㉓	㉔	㉕	㉖	㉗	㉘	㉙	㉚	㉛	㉜	㉝	㉞
MATERIAL																																		
COAL OR CHAR TONS/HR	977.25	364.37	564.88		2.12	1.00	7.88	1364.37		2177.50	964.42		59.22					564.88		962.42	2177.50						501.34	2.00	921.34	2.00					
C_n ②				88.97		190.78	190.78										0.06		0.18				0.31	0.25										
N_2		0.16		0.09	1.99		2.20	2.20	0.08	0.08		0.18	1.73		156.90	156.90																132.61	132.61
O_2															276.22		336.88	0.75		2.97		6068	6068	9.07	3.72	276.22								
H_2O		2.62		0.75	243.33		274.20	274.20	1.38	1.38		2.87	220.01					7.88		30.95				57.89	36.83									
CO_2		27.36		7.88	340.43		378.56	378.56	14.06	14.06		30.95	383.92					2.73		10.71				18.31	13.44									
CO		9.48		2.73	118.81		131.02	131.02	4.87	4.87		10.71	94.26					0.33		1.28				2.19	1.61									
H_2		1.13		0.33	14.07		15.67	15.67	0.56	0.56		1.28	11.04					1.08		4.26				7.27	5.34									
CH_4		3.77		1.08	43.39		52.07	52.07	1.93	1.93		4.26	13.81					0.05		0.19				0.33	0.24									
C_2H_6		0.17		0.05	1.71		2.37	2.37	0.09	0.09		0.19						0.13		0.51				0.87	0.64									
C_3H_8		0.65		0.13	4.46		6.20	6.20	0.23	0.23		0.51						0.05		0.18				0.21	0.15									
C_4H_{10}		0.16		0.05	0.94		2.18	2.18	0.08	0.08		0.18						0.03		0.12				0.60	0.44									
C_5H_{12}		0.11		0.03	0.66		1.52	1.52	0.06	0.06		0.12						0.09		0.35				2.36	1.88									
C_6^+ ③		0.31		0.09	1.86		4.28	4.28	0.16	0.16		0.35						0.38		1.50														
H_2S		1.52		0.58	9.39		18.30	18.30	0.66	0.66		1.50																						
1-95# STAGE TRANSPORT GAS																												144.24	144.24	92.07	52.17			144.24
TOTAL TONS/HR	1024.31	364.37	578.21	87.21	2.12	1089.35	1087.23	1408.54	24.17	2177.50	1076.2	624.55	59.22	432.72	156.90	336.85	13.55	564.66	55.20	962.42	2238.13	6063	9092	66.75	276.22	501.34	2.00	921.34	164.24	144.24	92.07	52.17	132.61	132.61	144.24
TEMP - °F	550	850	1540	1050	850	850	850	850	1000	1050	1540	1550	1550	1000	1000	1000	1000	1550	1000	1550	1040	1000	1000	1000	1000	800	800	800	800	800	195	195	100	486	195
PRESS - PSIA	2364		24.16	2364		20.20	19.84	27.06	27.06		26.08	27.06		30.52	30.52	30.52	24.16		26.08		30.52	30.52	29.48	28.18	30.52				21.56	21.20	2.50	2.50	300	600	270
H_2O MOL %	23.76		23.76	24.13		26.18	26.18	23.76	23.76		23.76	2.98		21.40	32.00	98.02	23.76		23.76		18.02	18.02	23.76	23.76	18.02				28.10	28.10	38.10	28.10		18.02	28.10
A/G MW	3.96		1.16	72.21		82.44	82.44	2.03	2.03		4.48	36.83		40.44	9.78	37.39	1.16		4.48		6.73	6.73	7.65	3.82	30.96				10.27	10.27	6.96	3.71		14.72	10.27
# MOL / HR																																0.33		4.13	44.58
# GPM			16.86	826.49		956.70	973.04	17.62	19.56		61.39	754.03		345.81	83.63	319.73	12.33		44.81		99.17	57.36	68.62	52.04	262.19				107.79	109.07	30.69	17.38		80.09	64.92
# ACFM	30.24		7.21	436.61		521.30	521.30	12.84	12.84		28.32	399.36		285.71	61.84	236.43	7.21		28.32		42.56	42.56	42.37	39.58	193.87				64.92	64.92	41.44	23.48		59.09	64.92
# SCFM ④	25.04																																	

① ASSUMED OIL MW OF 300 IN CALCULATING AVG MW
② AS n-C_6H_{14}
③ SCFM 60°F, 14.7 PSIA
④ EQUIPMENT SIZES INDICATED ARE FOR EACH TRAIN
⑤ FLOW RATES SHOWN ARE FOR TOTAL PLANT (BOTH TRAINS)

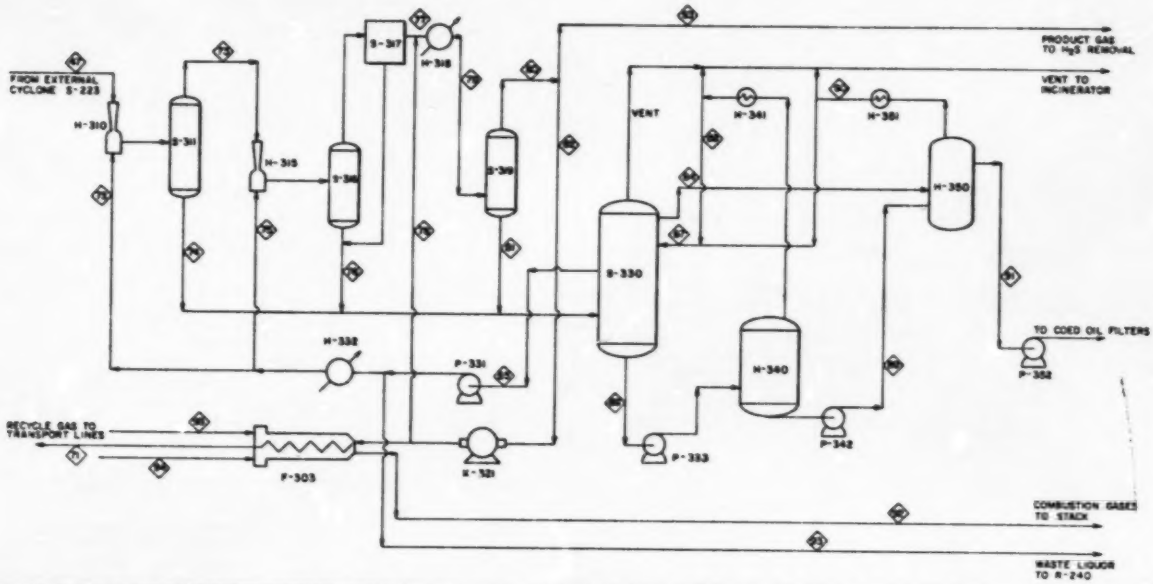
APPENDIX

Figure 3 (15)

PROCESS COED 24 M TON PER DAY PLANT
PRODUCT RECOVERY SYSTEM

FLOW SCHEME I
DUAL TRAIN PLANT

H-310 VENTURI SCRUBBER 487 MACFM	S-317 ELECTROSTATIC PRECIPITATOR 277 MACFM	S-329 GAS-WATER SEPARATOR 15' ϕ x 30'	P-333 DECANTER OIL PUMP 10 HP	H-350 OIL DEHYDRATOR 12' ϕ x 12'
S-311 GAS LIQUID SEPARATOR 22' ϕ x 40'	H-318 GAS COOLER 337 MM BTU/HR	P-331 RECYCLE LIQUOR PUMP 130 HP	H-340 HEAVY OIL DEHYDRATOR 12' ϕ x 12'	H-351 LIGHT OIL VENT CONDENSER 7 MM BTU/HR
H-315 VENTURI SCRUBBER 265 MACFM	H-332 SCRUB LIQUOR COOLER 218 MM BTU/HR	K-321 RECYCLE GAS COMPRESSOR 2500 HP	H-341 HEAVY OIL VENT CONDENSER 34 MM BTU/HR	P-352 OIL PUMP 10 HP
S-316 GAS LIQUID SEPARATOR 12' ϕ x 21'	F-303 RECYCLE GAS HEATER 44 MM BTU/HR	S-330 OIL WATER DECANTER 40' ϕ x 48'	P-342 DEHYDRATED OIL PUMP 10 HP	



STREAM NO / MATERIAL	69	70	70a	72	74	75	76	77	78	79	80	81	82	83	84	85	86	87	88	89	90	91	92	93	94	95
COAL OR CHAR TONS/HR	7.88			1.58	6.30		1.58								1.34		6.54			6.54		7.88				
OIL ①	190.78			38.16	152.62		35.46	2.70		2.70		2.70			32.43		158.35			158.35		190.78				
N_2	2.20	0.47		2.20				2.20	0.11	2.31	2.31		0.58	1.73									33.73			33.73
O_2																							0.49			10.18
H_2O	274.20	7.69	97.85	337.53	8721.87	8.30	8.30	337.53	1.77	339.30	38.04	301.26	9.46	28.58	7.48	685.94	31.67	38.16	31.67		8.49	0.99	6.64	236.94		1.19
CO_2	378.56	80.27		378.56				378.56	18.44	397.01	397.01		98.71	298.29									6.67			
CO	131.02	27.79		131.02				131.02	6.37	137.38	137.38		34.16	103.23												
H_2	15.67	3.32		15.87				15.67	0.76	16.43	16.43		4.08	12.35												
CH_4	52.07	11.04		52.07				52.07	2.53	54.60	54.60		13.57	41.03												
C_2H_4	2.37	0.50		2.37				2.37	0.12	2.49	2.49		0.62	1.87											2.43	
C_2H_6	6.20	1.32		6.20				6.20	0.30	6.50	6.50		1.62	4.88												
C_3H_6	2.18	0.47		2.18				2.18	0.12	2.30	2.30		0.59	1.71												
C_3H_8	1.52	0.32		1.52				1.52	0.07	1.59	1.59		0.39	1.20												
C_4 ⑫	4.28	0.91		4.28				4.28	0.21	4.49	4.49		1.12	3.37												
H_2S	18.30	3.88		18.30				18.30	0.89	19.19	19.19		4.77	14.42												
TOTAL TONS/HR	1087.23	137.98	97.85	991.64	8880.79	8.30	86.74	954.60	31.69	986.29	862.33	303.96	169.67	542.66	41.25	685.94	196.56	38.16	31.67	164.89	8.49	199.65	47.53	236.94	2.43	49.10
TEMP - °F	850	1065	150	190	190	150	165	195	280	110	110	110	110	110	183	183	183	183	183	250	183	250	1300	183	77	200
PRESS - PSIA	19.24	30.38		19.34				18.64	31.40	18.10	17.80		17.90	16.00									15.00		17.00	24.50
AVG MW	26.18	23.76		22.48				21.64	23.76	21.71	23.76		23.78	[illegible]									27.32		16.04	28.36
M MOL/HR	82.44	11.61		88.09				88.21	2.67	90.88	97.42		[illegible]	43.4									3.48		0.30	3.18
M GPM			23.59		24.33	3.385	3.550					[illegible]			0.184	28.26	0.784	0.157	0.131	0.692	0.027	0.790		0.975		
M ACFM	97304	99.74		529.23				954.09	11.25	311.59	328.66		83.14	274.72									72.99		1.68	19.31
M SCFM ③	821.30	75.41		567.02				937.78	10.88	974.66	363.09		90.20	272.79									22.01		1.90	20.11

① ASSUMED OIL MW OF 300 IN CALCULATING AVG MW
② AS n-C_4H_{10}
③ SCFM - 60°F, 14.7 PSIA
④ EQUIPMENT SIZES INDICATED ARE FOR EACH TRAIN
⑤ FLOW RATES SHOWN ARE FOR TOTAL PLANT (BOTH TRAINS)

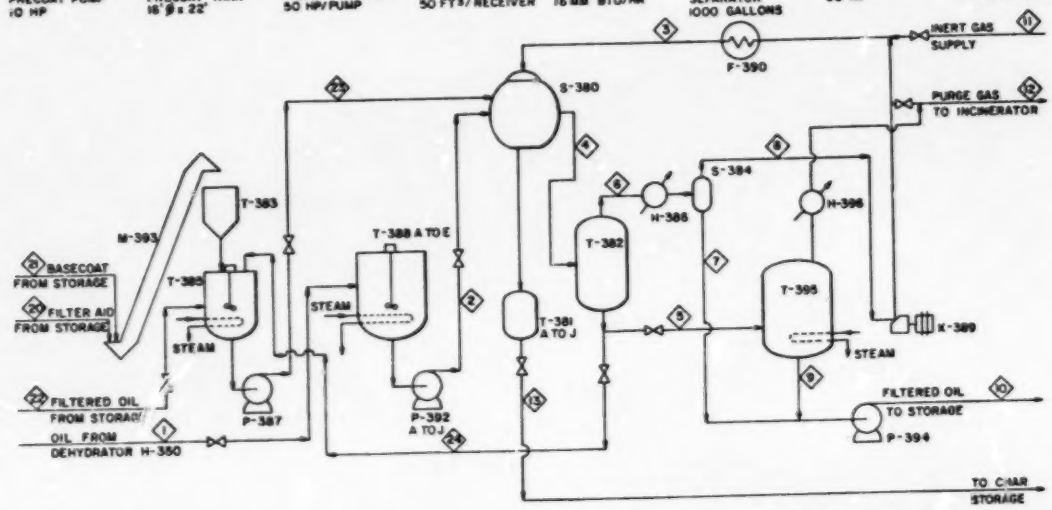
APPENDIX

Figure 4 (15)

PROCESS COED 24M TON PER DAY PLANT
COED OIL FILTRATION

SINGLE TRAIN PLANT EXCEPT AS NOTED

M-393 BUCKET ELEVATOR 8 TONS/HR
T-383 PRECOAT MIX BIN 15'ϕ x 25'
T-388 A TO E FILTER FEED TANK 5000 GALS/TANK
S-380 A TO J ROTARY PRESSURE PRECOAT FILTER 700 FT²/FILTER
T-382 FILTRATE RECEIVER 7300 GALLONS
F-390 PRESSURIZING-GAS PREHEATER 1.3 MM BTU/HR
T-395 FILTRATE HOLD TANK - 25,000 GAL
K-389 RECYCLE GAS COMPRESSOR 1000 HP
P-387 PRECOAT PUMP 10 HP
T-385 PRECOAT TANK 16'ϕ x 22'
P-392 A TO J FILTER FEED PUMP 50 HP/PUMP
T-381 A TO J SOLIDS RECEIVER 50 FT³/RECEIVER
H-386 LIGHT OIL CONDENSER 16 MM BTU/HR
S-384 LIGHT OIL CONDENSATE SEPARATOR 1000 GALLONS
P-394 FILTERED OIL PRODUCT PUMP 30 HP
H-396 LIGHT OIL CONDENSER 0.5 MM BTU/HR



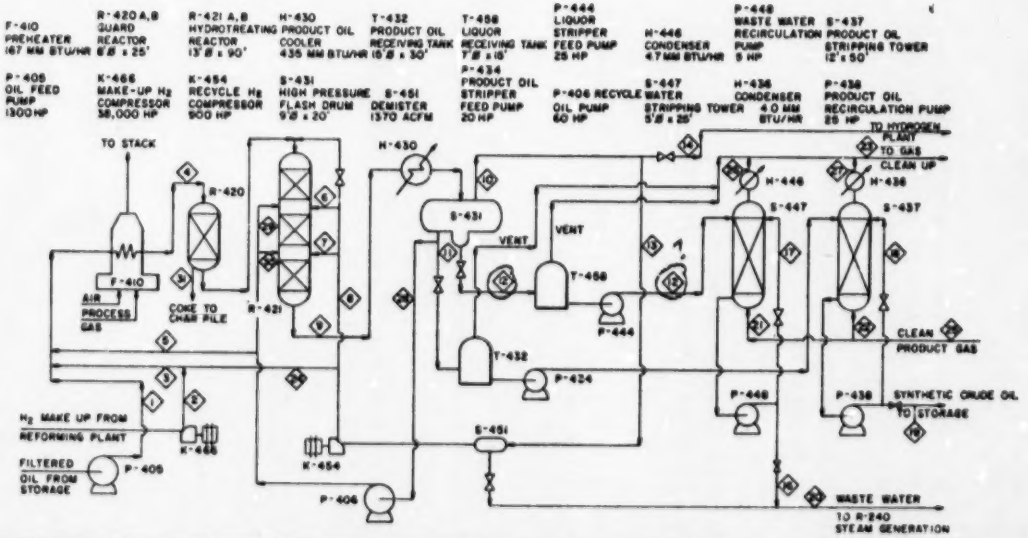
	CONTINUOUS FILTRATION (TONS/HOUR)														PRECOAT CYCLE (TONS/CYCLE) FILTRATION CYCLE LENGTH = 41 HR. PRECOAT CYCLE LENGTH = 7 HR.					
STREAM NO / MATERIAL	1	2	3	4	5	6	7	8	9	10	11	12	13		20	21	22	23	24	
OIL TONS/HR	190.78	190.78		184.96	175.71	9.25	9.25		175.71	184.96			5.82			FIRST PRECOAT ONLY	210.2	210.2	OFF SITE PRODUCT RECIRCULATION	
CHAR	7.88	7.88											7.88							
N_2			55.85	55.85		55.85		55.85			0.554	0.554								
H_2O	0.99	0.99	0.45	1.43		1.43	0.978	0.445		0.976		0.0045								
CO												TRACE								
CO_2												TRACE								
CH_4												TRACE								
C_2^+												TRACE								
H_2S																				
FILTERAID													1.52		63.1			63.1		
BASECOAT MATERIAL																0.068				
TOTAL TONS/HR	199.65	199.65	56.30	242.24	175.71	66.53	10.23	56.30	175.71	185.94	0.554	0.558	15.22		63.1	0.068	210.2	273.3		
TEMP °F	250	340	400	350	350	350	80	80	330	300	70	100	350				250	320		
PRESSURE PSIA			60	42	42	42	41	41			60	60								
AVE. M.W.			27.88	27.62		27.62		27.88			28.0	27.83								
MOL/HR			40395	4148		4148		4039			39.6	40.1								
GPM	790	790		673	639		39		639	673										
ACFM			10300	14300		14300		9490			62	67								
SCFM			25500	26200		26200		25500			250	253								

APPENDIX

Figure 5 (15)

PROCESS COED 24M TON PER DAY PLANT

HYDROTREATING PLANT



MATERIAL \ STREAM NO	①	②	③	④	⑤	⑥	⑦	⑧	⑨	⑩	⑪	⑫	⑬	⑭	⑮	⑯	⑰	⑱	⑲	⑳	㉑	㉒	㉓	㉔	㉕	㉖	㉗	㉘	㉙	㉚	㉛	㉜
COKE TONS/HR																																
OIL	184 96			228 86	43 90				296 10		164 40							82 20	164 40													0 04
H_2		3 50	5 39	5 39		190	1 90	3 80	9 19	9 19			5 69	3 50							0 18	0 70	0 88	1 89	0 88	0 18	0 70				131 70	43 90	43 90	
H_2O	0 98		0 04	1 02		0 03	0 03	0 06	17 99	0 16		17 83	0 10	0 06	16 61	16 61			16 61		0 22	0 88	2 30	0 04	1 08	1 44	0 88						
CO_2		0 46	0 71	0 71		0 25	0 25	0 50	1 21	1 21			0 75	0 46							1 89	7 55	9 44	0 25	9 44	1 89	7 55						
CO		3 50	5 65	5 65		2 14	2 14	4 28	10 38	10 38			6 43	3 95							16 01	63 99	79 90	2 15	79 90	16 01	63 99						
H_2		20 39	26 36	26 36		5 97	5 97	11 94	28 92	28 92			17 91	11 01							1 02	4 08	5 10	5 97	5 10	1 02	4 08						
CH_4		0 50	1 72	1 72		1 23	1 23	2 46	5 94	5 94			3 68	2 26							1 27	5 09	6 36	1 22	6 36	1 27	5 09						
C_2H_4			0 03	0 03		0 04	0 04	0 08	0 18	0 18			0 11	0 07										0 03								
C_2H_6			0 71	0 71		0 72	0 72	1 44	3 47	3 47			2 19	1 32										0 71								
C_3H_8			0 14	0 14		0 14	0 14	0 28	0 68	0 68			0 42	0 26										0 14								
C_4H_{10}			0 97	0 97		0 56	0 56	1 12	2 73	2 73			1 69	1 04										0 97								
C_6+ ①			0 88	0 88		0 87	0 87	1 74	4 23	4 23			2 62	1 61										0 88								
H_2S			1 86	1 86	0 02	1 85	1 85	3 70	9 74	8 96	0 07	0 63	5 56	3 42							0 01	0 03	0 74	1 86	0 04	0 64	0 10	0 06	0 02	0 02		
NH_3			0 06	0 09	0 03	0 06	0 06	0 12	2 63	0 29	0 11	2 14	0 18	0 11									2 25	0 06		2 14	0 11	0 09	0 03	0 03		
TOTAL TONS/HR	185 94	28 35	44 12	274 01	43 95	15 76	15 76	31 52	393 39	76 36	164 58	20 60	47 29	29 07	16 61	16 61	82 20	164 40	16 61	20 60	82 20	106 97	18 77	102 80	24 99	82 38	131 85	43 99	42 35	0 04		
TEMP °F	250	200	200	650	100	200	200	200	750	100	100	100	100	100		100	100	100	100	100	200	200	100	200	200	100	100	100	100	100	650	
PRESS - PSIA	1800	1800	1800	1720	1800	1800	1800	1800	1710	1700			1700	16							20	20	16	1800	20	17	17	1700	1800	1800		
AVG MW		2 72	3 22	3 28		4 77	4 77	4 77	5 67	4 77			4 77	4 77							16 99	16 99	16 97	4 77	16 99	17 21	16 90					
M MOL/HR	*	20 81	27 42	27 53		6 61	6 61	13 22	34 31	32 02			19 83	12 19							2 44	9 73	12 61	6 61	12 17	2 86	9 79					
M GPM	0 743				0 194						0 728	0 083																				
M ACFM		1 36	1 80	3 18		0 43	0 43	0 87	4 34	1 89			1 17	76 27														0 563	0 194	0 194		
M SCFM ②		131 59	173 39	174 06		41 80	41 80	83 59	216 95	202 47			125 36	77 08							14 39	57 40	78 89	0 43	71 79	16 85	57 41					
																					18 43	61 53	79 76	41 80	76 95	18 09	61 68					

① AS n-C_6H_{14}

② SCFM - 60°F, 147 PSIA

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